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Dual Alkali Acceptance Test at Louisville Gas and Electric Company; Volume I. Acceptance Test and Appendices A-C

Bechtel Corp.
San Francisco, CA

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

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DUAL ALKALI ACCEPTANCE TEST AT LOUISVILLE GAS AND ELECTRIC COMPANY; VOLUME I. ACCEPTANCE TEST AND APPENDICES A-C

Prepared for

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Office of Environmental Engineering and Technology

Prepared by

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16. ABSTRACT The report gives results of the completed acceptance test series run on the dual alkali system serving Louisville Gas and Electric Company's Cane Run Unit 6 boiler. Test results, measuring system performance, included: (1) SO2 removal averaged 94% and 143 ppm outlet concentration; (2) soda ash consumption averaged 0.042 mole soda ash per mole SO2 removed; (3) lime consumption averaged 1.04 mole CaO per mole SO2 removed; (4) power consumption averaged 1.05% of generation; (5) filter cake solids averaged 52.2 wt % insoluble solids; and (6) there was no net particulate matter addition. Various problems attributable to the boiler, the FGD system, and the quality and quantity of the carbide lime supplied to the system delayed acceptance testing until July 1980. The year-long demonstration period officially started in May 1980. The problems experienced and their solutions are discussed. This volume contains the process description and a discussion of the test results, operating history, and performance. Appendices A-C contain acceptance test daily summaries, monthly performance parameters, and Kenvirons' acceptance test reduced data. Volume II contains Appendices D-F, and Volume III contains Appendices G-J.

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Sulfur Dioxide	Pollution Control	13B	07B
Boilers	Sodium Carbonates	Stationary Sources	13A	
Flue Gases	Calcium Oxides	Dual Alkali System	21B	
Scrubbers	Dust	Particulate	07A, 13I	11G
Desulfurization	Aerosols		07D	
Tests	Acceptability		14B	14G

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EPA-600/7-81-159a

September 1981

DUAL ALKALI ACCEPTANCE TEST AT LOUISVILLE GAS AND ELECTRIC COMPANY

VOLUME I

Acceptance Test and Appendices A-C

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i.b

ABSTRACT

A dual alkali flue gas desulfurization (FGD) process developed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) was installed to control sulfur dioxide emissions from Unit 6, coal-fired boiler at Louisville Gas and Electric Company's (LG&E) Cane Run Station. The Federal Environmental Protection Agency (EPA) has selected this system as the demonstration plant for dual alkali technology and is participating in funding of the operation, testing and reporting of the project.

Bechtel National, Inc. (Bechtel) was contracted to provide an independent test program to characterize the system operation over a year-long demonstration and to evaluate the process with respect to its performance guarantees.

The overall program consists of four phases:

Phase I - Preliminary Design and Cost Estimation
Phase II - Engineering Design, Construction and Mechanical Testing
Phae III - Start-up and Acceptance Testing
Phase IV - One Year Demonstration

This is a joint report covering Phase III and part of Phase IV which include:

- Charging of the system
- Start-up of the process
- First four months of Demonstration
- 10-day Acceptance Test

The results of the Acceptance Test were as follows:

- SO_2 removal averaged 94% with 143 ppm outlet SO_2 concentration
- Soda ash consumption averaged 0.042 mole Na_2CO_3 per mole SO_2 removed
- Lime consumption averaged 1.04 mole CaO per mole SO_2 removed
- Power consumption averaged 1.05% of generation
- Filter cake solids averaged 52.2 wt% insoluble solids
- There was no net particulate matter addition

ACKNOWLEDGMENTS

The presentation of this report was conducted under the direction of N. Kaplan, EPA Project Officer, Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina. This joint report covers the Acceptance Test conducted by Bechtel National, Inc. under EPA Contract No. 68-02-2634 and the Phase 3 operation of the dual alkali system by Louisville Gas and Electric Company under EPA Contract No. 68-02-2189.

Input for the report was supplied by L. R. Woodland and C. R. La Mantia of Arthur D. Little, Inc., 20 Acorn Park, Cambridge, MA 02140 and R. C. Tussey of Kenvirons, Inc., 452 Versailles Road, Frankfort, KY 40602. Critical review was supplied by R. P. Van Ness of Louisville Gas and Electric Company, 311 W. Chestnut Street, Louisville, KY 40201.

Contributing to the organization, technical input, and review of the report were N. Zoueshtiagh, T. M. Martin, and J. K. Donnelly of Bechtel National, Inc., L. Ylagan provided typing of the manuscript.

PHASE III START-UP AND ACCEPTANCE TESTING

PHASE IV DEMONSTRATION TESTING (INTERIM)

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NOTES

1. Company Names and Products

   The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Units of Measure

   EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL-RTP provides conversion factors for the non-metric units. Generally, this paper uses British units of measure.

   The following equivalents can be used for conversion to the Metric System:

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 m
1 ft^2	0.0929 m^2
1 ft^3	0.0283 m^3
1 grain	0.0648 gram
1 in.	2.54 cm
1 $in.^2$	6.452 cm^2
1 $in.^3$	16.39 cm^3
1 lb (avoir.)	0.4536 kg
1 ton (long)	1.0160 m tons
1 ton (short)	0.9072 m tons
1 gal.	3.7854 liters
1 Btu	252 calories

Section 1

INTRODUCTION

1.1 Purpose of Program

The purpose of the program is to demonstrate the full-scale utility application of dual alkali (D/A) flue gas desulfurization (FGD) technology. It is a joint effort by a number of organizations under the sponsorship of the Environmental Protection Agency (EPA). The system was designed by Combustion Equipment Associates, Inc. (CEA)*/Arthur D. Little, Inc. (ADL) and installed on Unit 6, an existing 300 MW (gross peak capacity) coal-fired boiler at Louisville Gas and Electric Company's (LG&E) Cane Run Station in Louisville, Kentucky. Installation was done by the construction department of LG&E under the supervision of CEA/ADL.

The FGD system was installed to comply with the requirements of the Jefferson County Air Pollution Control District, the Kentucky State Division of Air Pollution and EPA Region IV (85% removal of SO_2). EPA selected the D/A process at LG&E as the demonstration system for this technology and is participating in the funding of the operation, testing, and reporting of the project.

Bechtel National, Inc. (Bechtel) is under separate contract with EPA to provide an independent test program to assess the performance of the system with respect to the guarantees offered LG&E by CEA/ADL and to

*After Chapter 11 bankruptcy proceedings, the CEA air pollution control group has been taken over and is currently operating as part of Thyssen Environmental Systems, Inc. ADL is now under direct contract to LG&E.

monitor and characterize the performance of the system over a year-long demonstration period. Kenvirons, Inc. is the test subcontractor to Bechtel providing stack testing services for the program.

1.2 Scope of Work

The scope of work for the program includes the design, construction, start-up, Acceptance Testing, and one year of demonstration of the CEA/ADL concentrated-mode dual alkali process. The system is designed to treat all of the flue gas at the nominal rated capacity (280 MW) with the capability of treating flue gas equivalent to a minimum boiler load of 60 MW and a maximum load of 300 MW.

LG&E is the prime contractor with overall responsibility for the construction and operation of the system. CEA, under contract to LG&E, was responsible for the engineering design, for the supply of all process equipment, and for engineering assistance during start-up and Acceptance Testing. CEA was also responsible for compliance with all process guarantees and equipment warranties. ADL, a subcontractor to CEA, provided process engineering support to CEA during design, start-up, and Acceptance Testing; and will provide process assistance to LG&E in the operation of the system during the one year test program. ADL is also responsible for the preparation of all reports required under the EPA/LG&E contract.

The overall program is divided into four phases:

- Phase I - Preliminary design and cost estimates
- Phase II - Engineering design, construction, and mechanical testing
- Phase III - Start-up and Acceptance Testing

- Phase IV - One year Demonstration

Bechtel is the test contractor with overall responsibility for the design and execution of the Demonstration Test program. Through the field laboratory established on site, the operation of the system will be closely monitored over the demonstration year. The test program includes examination of:

- SO_2 removal
- Lime utilization
- Soda ash makeup
- Regeneration of spent liquor
- Water balance
- Materials of construction
- Operating problems
- Waste cake properties and disposal
- Reliability and availability
- Economics (capital investment and operating cost)

This report covers work performed in Phase III and part of Phase IV of the program in which the following activities were completed.

- Charging of the system
- Start-up of the process
- First four month of the year-long Demonstration
- 10-day Acceptance Test

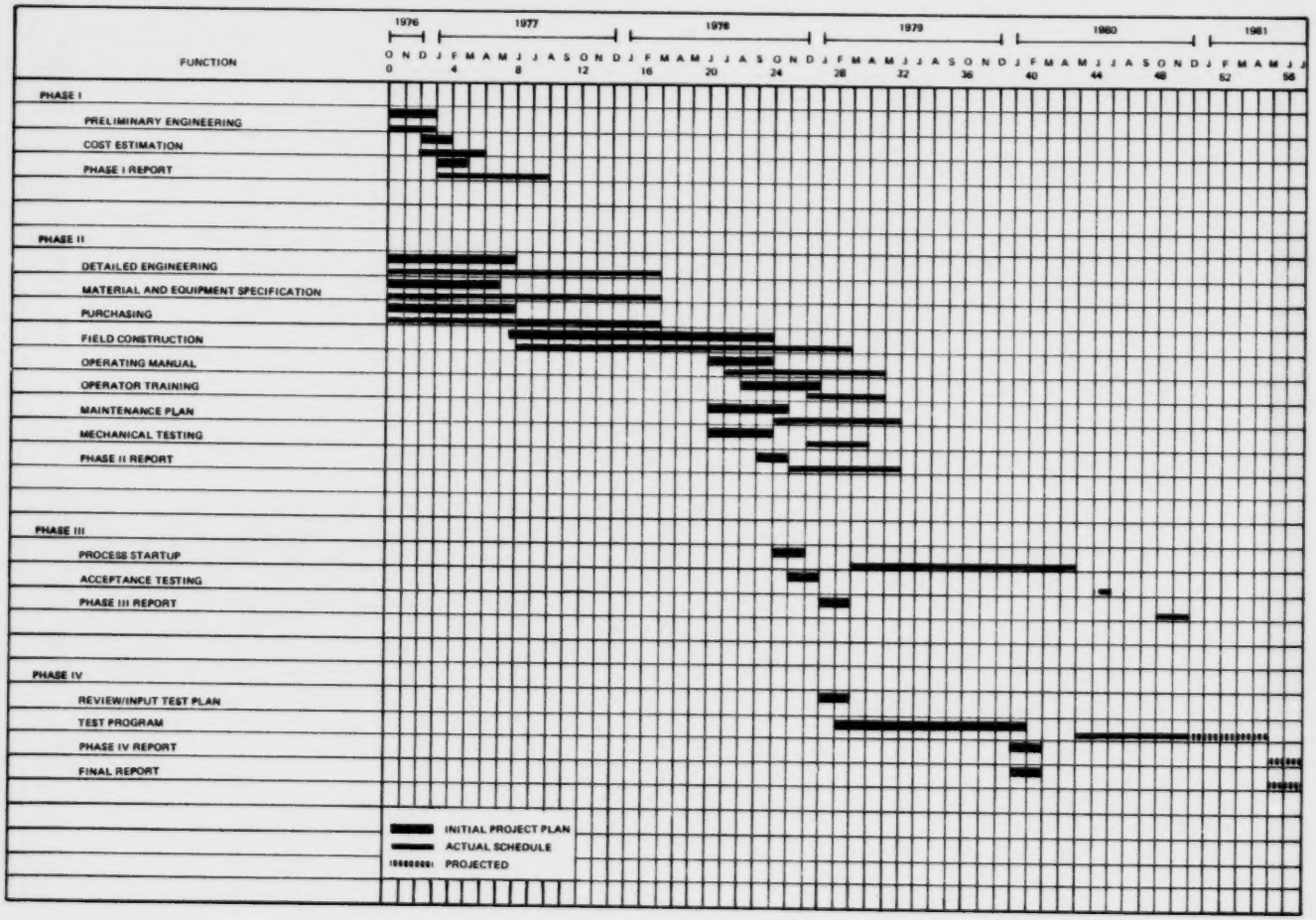
The report contains a detailed description of the plant, the planned mode of operation, the operating history and the results.

1.3 Program Schedule

The overall project schedule covering all phases of the dual alkali demonstration program, is given in Figure 1.1. The overall program, including the one year demonstration was originally scheduled for 40

FIGURE 1.1

DUAL ALKALI DEMONSTRATION PROJECT OVERALL SCHEDULE



months (with an additional month for completion of the final draft report).

Phase I and II were scheduled to begin simultaneously to expedite the overall project. Phase I (preliminary design) was scheduled to be completed in five months including preparation of the draft report. Phase II (engineering design and construction) was scheduled to be completed 24 months from the signing of the contract.

Acceptance Testing was originally scheduled to precede the year-long Demonstration. However, it was delayed on six occasions for lack of readiness. Because of repeated delays, with agreement of LG&E, on April 30, 1980, the year-long Demonstration run went into effect. Six weeks later, between July 17 and 28 the Acceptance Test did take place.

Section 3 of this report addresses the initial operation of the system and a portion of the Demonstration program.

Section 4 addresses the Acceptance Test itself.

Section 2

CEA/ADL DUAL ALKALI PROCESS AND THE FULL-SCALE LG&E SYSTEM

2.1 General Description of Dual Alkali Process

Dual alkali (or double alkali) is a generic term used to describe flue gas desulfurization (FGD) systems involving the absorption of SO_2 by a soluble alkali, followed by reaction of the spent scrubber solution with lime and/or limestone to regenerate the alkali and produce a waste calcium-sulfur salt for disposal.

The two principal features of dual alkali technology which set it apart from conventional direct lime and limestone scrubbing are: (1) the use of a clear solution rather than a slurry for contacting the flue gas in the absorber; and (2) reaction of the solution in a separate absorbent regeneration section to form the waste solids rather than forming the waste solids as a part of the scrubbing operation. The use of scrubbing solution (rather than slurry) has a number of important advantages. First, with alkaline solutions, high SO_2 removal efficiencies can be achieved over a wide range of inlet SO_2 concentrations. Second, precipitation of the waste calcium sulfite/sulfate solids outside the scrubber facilitates formation of waste solids with good dewatering properties. There is minimal scale potential in the scrubber circuit.

A process flow diagram for a generalized dual alkali system using commercial grade lime is shown in Figure 2.1.

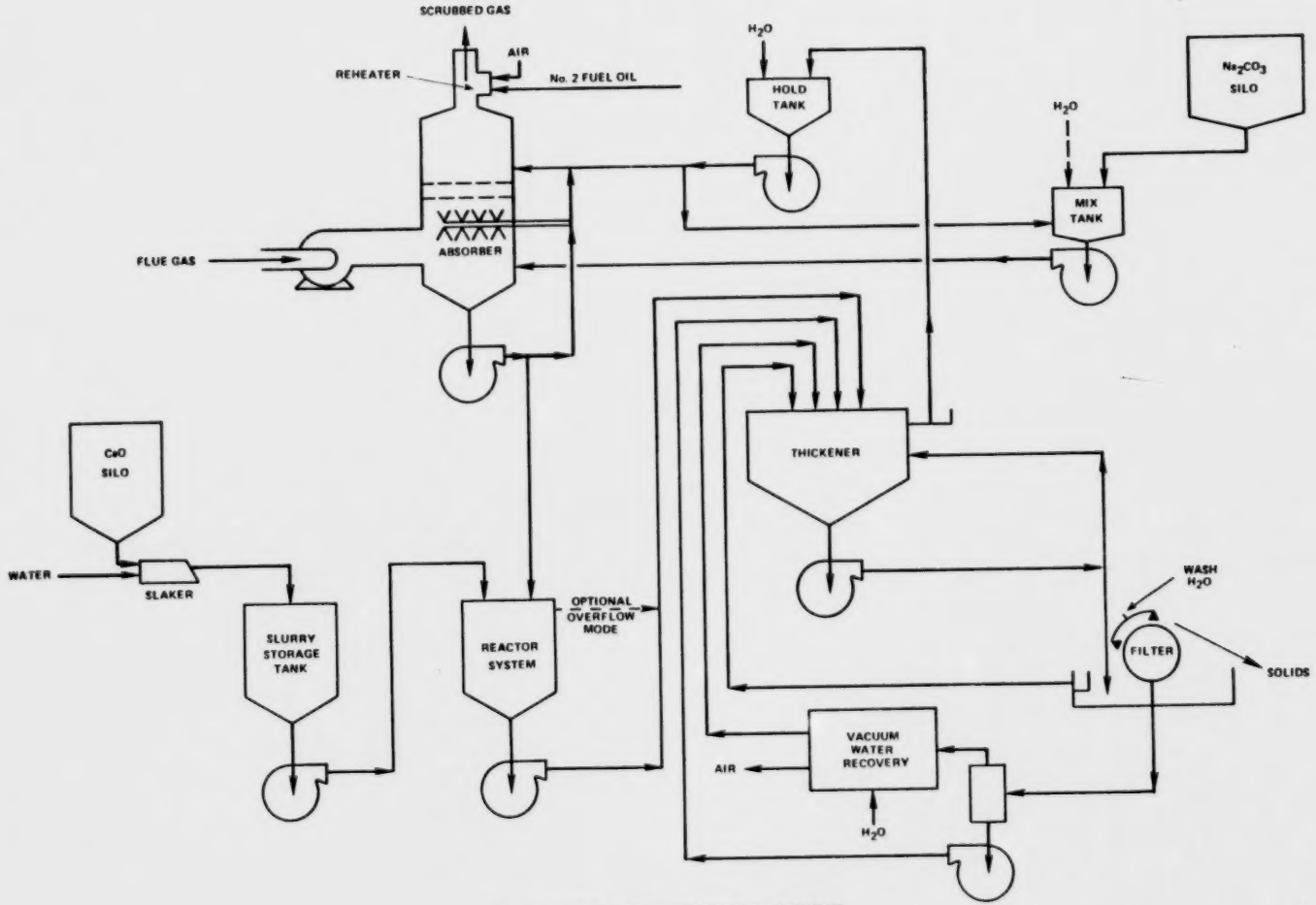


Figure 2–1 DUAL ALKALI PROCESS FLOW DIAGRAM

2.2 CEA/ADL Dual Alkali Process

The CEA/ADL process is a concentrated-mode dual alkali system in which the active sodium concentration is maintained at 0.4 - 0.5 M. The primary active sodium species in the SO_2 absorption step is sodium sulfite, but oxidation of sodium sulfite/bisulfite also generates a high concentration ($\simeq$1 M) of the inactive sodium sulfate. In regeneration, the sulfate (as calcium sulfate) is occluded and purged with the predominantly sulfite solids.

Based on the major steps of the dual alkali technology, the flue gas desulfurization system can be broken down into five sections: gas scrubbing, regeneration of scrubbing solution, solids separation, raw materials preparation, and waste disposal.

2.2.1 Flue Gas Scrubbing

In the gas scrubbing section, SO_2 is removed by contacting the alkaline scrubbing solution with flue gas on the trays of the absorber. The major "active" constituent of the scrubbing liquor, sodium sulfite, reacts with SO_2 in the flue gas to form bisulfite as follows:

$$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3 \qquad (1)$$

There may also be small amounts of the hydroxide and carbonate present in the regenerated scrubbing liquor from the thickener. To reduce the potential for (calcium) carbonate scale formation on the trays, a portion of the slightly acidic liquor from the bottom of the absorber is blended with the alkaline regenerated liquor from the thickener to

maintain a pH of below 10. This slightly acidic recirculated liquor reacts with the carbonate and hydroxide to yield additional sulfite:

$$Na_2CO_3 + 2NaHSO_3 \longrightarrow 2Na_2SO_3 + H_2O + CO_2 \quad (2)$$

$$NaOH + NaHSO_2 \longrightarrow Na_2SO_3 + H_2O \quad (3)$$

Oxidation of the sodium sulfite to sodium sulfate:

$$2Na_2SO_3 + O_2 \longrightarrow 2Na_2SO_4 \quad (4)$$

must be held to less than 20% by limiting (1) oxygen concentration in the flue gas or (2) number of absorber modules in operation at low load.

Otherwise the rate at which sulfate can be purged as a mixed crystal will be increased and the solids settling rate will be deteriorate.

2.2.2 Absorbent Regeneration

The active sodium fed to the absorber is defined as:

$$[\text{active sodium}] = 2 \times [Na_2SO_3] + [NaOH] + 2 \times [Na_2CO_3] + [NaHSO_3] + [NaHCO_3]$$

The activity is restored by reaction with lime slurry in a two-step process:

$$Ca(OH)_2 + 2NaHSO_3 \longrightarrow CaSO_3.1/2\ H_2O\downarrow + Na_2SO_3 + 3/2\ H_2O \quad (5)$$

$$Ca(OH)_2 + Na_2SO_3 + 1/\ 2H_2O \longrightarrow CaSO_3.1/2\ H_2O\downarrow + 2NaOH \quad (6)$$

The first reaction is a neutralization reaction which goes to completion. In the second, the equilibrium hydroxide concentration is limited by the relative solubility of calcium sulfite and calcium hydroxide and the concentrations of hydroxide. Both of these reactions lead to precipitation of calcium sulfite hemihydrate.

A limited amount of calcium sulfate is also precipitated:

$$Ca^{++} + SO_4^{=} \longrightarrow CaSO_4\downarrow \qquad (7)$$

In a concentrated-mode, the sulfate co-precipitates with the calcium sulfite and gypsum, the dihydrate is not formed, allowing system to operate unsaturated with respect to calcium sulfate.

The regeneration is accomplished in a two-stage reactor system to produce solids with improved settling and filtering characteristics.

2.2.3 Solids Dewatering

Solids dewatering is a unit operation involving thickening of the reactor effluent slurry to 15-30 wt % solids, followed by filtration to produce a waste filter cake. While the filter cake is being formed, it is washed with fresh water to recover sodium salts; otherwise sodium is lost in the process liquor discharged with the cake. The wet filter cake constitutes the only waste discharged from the process. There is no other purge from the system.

The solids content and chemical composition of the waste cake produced from SO_2 control depend primarily upon the amount of oxidation, the quality of lime, and the chloride content of the coal. For utility boilers firing medium to high sulfur coal, the waste cake for a system operating on commercial lime would contain 55-70 wt % insoluble solids and have a chemical composition typically within the following ranges:

Species	Weight % (Dry Basis)
$CaSO_3 \cdot 1/2H_2O$	80 - 85
$CaSO_4$	10 - 15
$CaCO_3$ + Inerts	5 - 10
Soluble Sodium Salts	1 - 2

The sodium lost in the cake will depend on the sodium concentration in the process liquor, the amount of oxidation, the chloride content of the coal, and the permissible level of cake washing.

2.2.4 Raw Materials Preparation

Two chemicals are required for the operation of the dual alkali system; lime for regeneration and soda ash to make up for sodium losses. Typically, lime would be hydrated in a slaker prior to being fed to the reactor system. At the Cane Run Station, however, carbide lime is available. Carbide lime is a well-hydrated calcium hydroxide slurry containing 25 - 30% insoluble salts.

The slurry is pumped to a day-tank from which it is fed to the primary reactors.

The rate of addition of soda ash to the system is adjusted to compensate for the losses of sodium in the cake.

2.2.5 Waste Disposal

The FGD waste solids from the filters must be treated to produce a more physically stable and water impermeable sludge for disposal. One such treatment method, the IU Conversion System process (in which filter cake

is blended with fly ash and lime) is being used at Cane Run. This waste treatment/disposal system is described in more detail in Section 2.3.

2.3 Process Design LG&E Full-Scale Dual Alkali Scrubber System

2.3.1 Unit 6 Boiler

Cane Run Unit 6 consists of a pulverized coal-fired steam boiler, built by Combustion Engineering, with a Westinghouse turbine-generator. The unit operates from a minimum of 60 MW during off-peak hours to a maximum of 300 MW during peak hours. The annual average load is equivalent to approximately 180 MW (about 60% of the gross peak capacity).

Flue gas from the boiler passes, in parallel streams, through two Ljungstrom combustion air preheaters. Each preheater discharges flue gas through separate ducts to an electrostatic precipitator designed for 99.4% removal efficiency of particulate matter. From the precipitator, the gases enter two parallel induced draft fans, each handling 50% of the total gas.

The sulfur dioxide removal system, installed between the existing induced draft fans and the stack, draws hot flue gas from the outlet of the induced draft fans through two booster fans. The scrubbed gas is reheated with hot air from an oil-fired burner, then returned to the existing entrance to the stack. Dampers have been provided to allow bypass of gas around the scrubber system using the existing ductwork.

Coal for Unit 6 is received from a number of sources. The average sulfur

content on a dry basis is 3.5%. The design for the LG&E dual alkali system is based on 5% sulfur coal. The average chloride content of the coal is 0.04%, ranging from 0.03% to 0.06%.

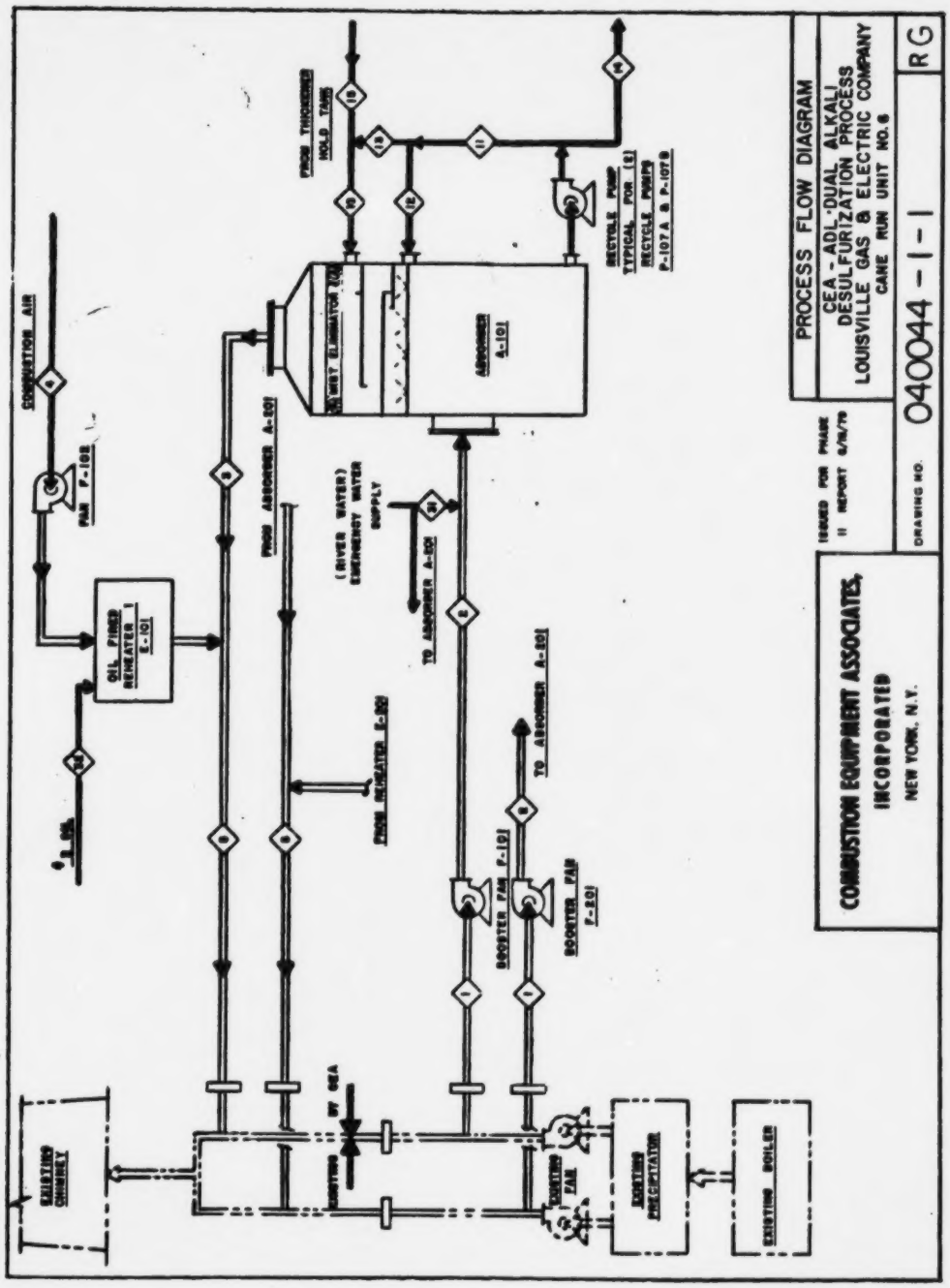
2.3.2 Dual Alkali FGD System

The dual alkali system for Unit 6 can be described in five parts: (1) absorber section (flue gas scrubbing); (2) reactor section (regeneration of scrubbing solution); (3) solids dewatering; (4) raw materials preparation; and (5) waste disposal. The process flow diagrams for the overall system are shown in Figures 2.2 and 2.3 (CEA Drawings 040044-1-1 and 040044-1-2). Tables 2.1 and 2.2 show the design material balance.

Absorber Section - The absorber section consists of two identical scrubber modules. Each module is made up of a booster fan, an absorber, an oil-fired flue gas reheater, and two recirculation pumps (one operating and one back-up).

Flue gas is drawn from the existing induced draft fans and is forced through the absorbers by means of the booster fans. The basic design and control philosophy is predicated on both absorbers operating simultaneously with each handling half the boiler load. However, during periods of low boiler load, one absorber is capable of handling all of the gas. Hence, in order to allow for greater flexibility in operation and provide for the possibility of maintenance on one module while the system is still in service, a common duct connecting the two booster fan inlets has been incorporated. A bypass is also provided to allow complete shutdown of the scrubbers while the boiler is still on-line. The bypass and FGD system

Figure 2.2


PROCESS FLOW DIAGRAM
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(RIVER WATER) EMERGENCY WATER SUPPLY
TO ABSORBER A-101
BOOSTER FAN P-101
BOOSTER FAN P-201
TO ABSORBER A-201
ABSORBER A-101
MIST ELIMINATOR
FROM THICKENER HOLD TANK
RECYCLE PUMP TYPICAL FOR (5) RECYCLE PUMPS P-107A & P-107B
EXISTING CHIMNEY
EXISTING FAN
EXISTING PRECIPITATOR
EXISTING BOILER

Figure 2.3

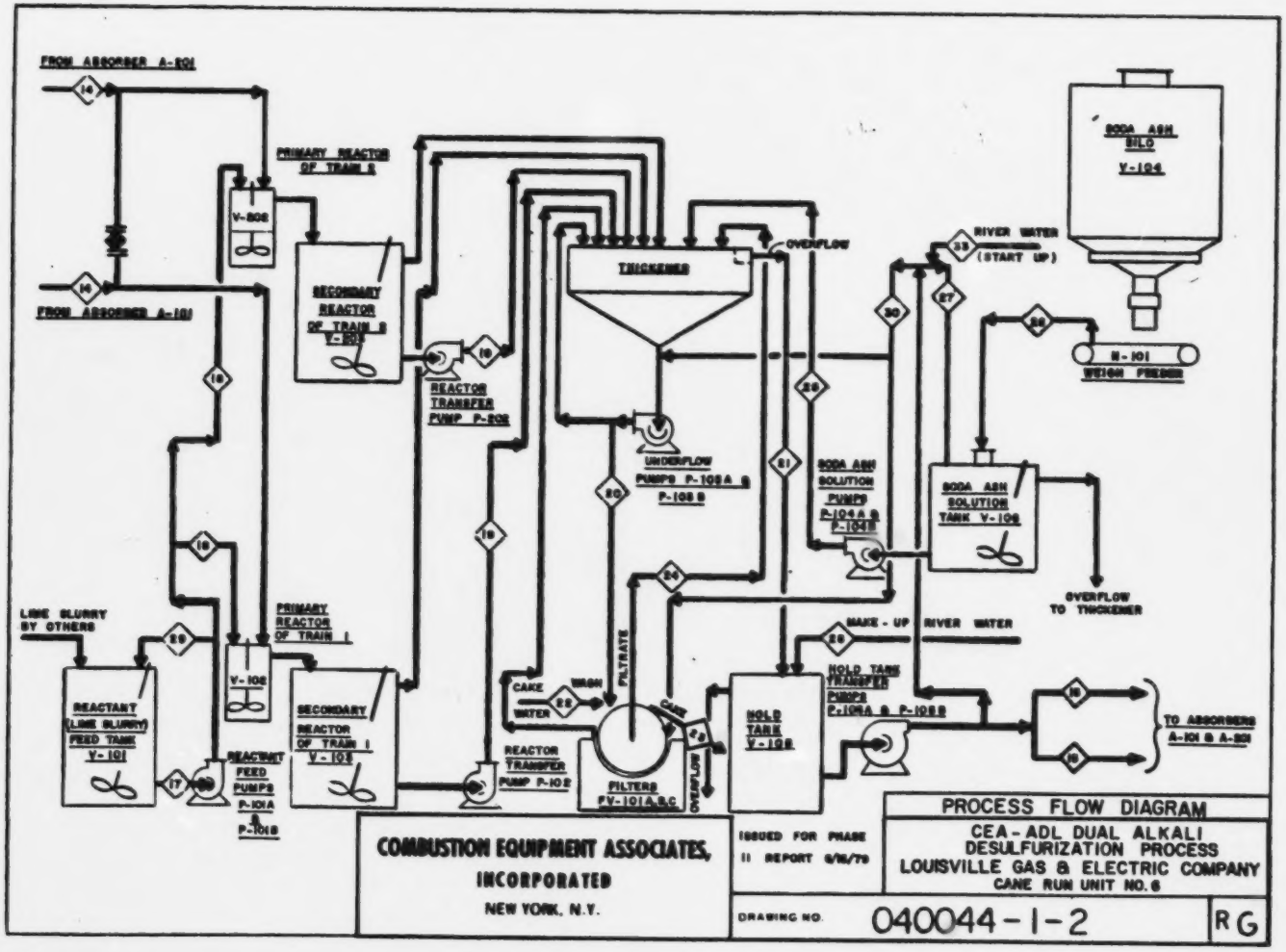
FROM ABSORBER A-101
PRIMARY REACTOR OF TRAIN II
V-102
SECONDARY REACTOR OF TRAIN II V-103
REACTOR TRANSFER PUMP P-102
THICKENER
OVERFLOW
SODA ASH SILO V-104
RIVER WATER (START UP)
W-101 WEIGH FEEDER
UNDERFLOW PUMPS P-103A & P-103B
SODA ASH SOLUTION PUMPS P-104A & P-104B
SODA ASH SOLUTION TANK V-105
OVERFLOW TO THICKENER
LIME SLURRY BY OTHERS
PRIMARY REACTOR OF TRAIN I
V-106
SECONDARY REACTOR OF TRAIN I V-107
REACTANT (LIME SLURRY) FEED TANK V-101
REACTANT FEED PUMPS P-101A & P-101B
REACTOR TRANSFER PUMP P-102
CAKE WATER
WASH
FILTRATE
CAKE
FILTERS FV-101A,B,C
OVERFLOW
MAKE-UP RIVER WATER
HOLD TANK V-108
HOLD TANK TRANSFER PUMPS P-108A & P-108B
TO ABSORBERS A-101 & A-201
COMBUSTION EQUIPMENT ASSOCIATES, INCORPORATED
NEW YORK, N.Y.
ISSUED FOR PHASE II REPORT 6/15/79
PROCESS FLOW DIAGRAM
CEA-ADL DUAL ALKALI DESULFURIZATION PROCESS
LOUISVILLE GAS & ELECTRIC COMPANY
CANE RUN UNIT NO. 6
DRAWING NO. 040044-1-2
RG

Table 2.1

MATERIAL BALANCE - ABSORBER SECTION*

	I.D. Fan Outlet	Absorber Inlet	Absorber Outlet	Combustion Air	Exit Flue Gas
Stream No.	1	2	3	4	5
Volume, ACFM	532,907	525,880	436,516	11,000	487,215
Temperature °F	300	303	126	60	176
Pressure, inches WG	+2.0	+11.5	+3.5	AMB	+2.0
Dry Gas, #/min	26,699	26,699	26,515		
H_2O Vapor, #/min	1,316	1,316	2,474		
Total Gas, #/min	28,015	28,015	28,989	850	27,839
SO_2, #/min	194.95	194.95	11.25		11.25
SO_2,PPM (Dry Vol)	3,471	3,471	200		200
Particulates, #/min	2.48	2.48	2.48		2.48

	Combined Recycle	Spray Recycle	Tray Recycle	Absorber Bleed	Feed Forward	Tray Feed	Emergency Spray Water	Reheater Oil
Stream No.	11	12	13	14	15	16	31	32
H_2O, #/min	19,933.5	12,919.5	7,014.0	13,642.8	14,828.1	21,867.0	4,582.0	
Na_2CO_3, #/min					6.8			
NaOH, #/min					50.0			
Na_2SO_3, #/min	174.9	113.4	61.5	119.7	347.2	582.0		
$NaHSO_3$, #/min	666.9	432.2	234.7	456.4		95.0		
Na_2SO_4, #/min	1,542.9	1,000.0	542.9	1,056.0	1,016.3	1,559.2		
$Ca(OH)_2$, #/min								
$CaSO_3 \cdot 1/2\ H_2O$, #/min								
$CaSO_4 \cdot 1/2\ H_2O$, #/min								
$CaCO_3$, #/min								
Inerts, #/min								
CaCl, #/min	445.8	288.9	156.9	305.1	303.6	460.5		
Total, #/min	22,764.0	14,754.0	8,010.0	15,580.0	16,552.0	24,563.7	4,582.0	
% Solids, #/min	0	0	0	0	0	0	0	
Flow, GPM	2,481.0	1,608.0	873.0	1,698.0	1,837.0	27,101.0	550.0**	3.0
pH	4-8	4-8	4-8	4-8	7-13	6-11		
Temp, °F	120-140	120-140	120-140	120-140	120-140	120-140		AMB

* For scrubber module

** Emergency only, Max. Flow

Combustion Equipment Assoc.
Material Balance
for
Drawing 040044-1-1 Rev G

Table 2.2

MATERIAL BALANCE - REACTOR, SOLIDS DEWATERING AND RAW MATERIALS SECTIONS

Stream No.	Absorber Bleed 14*	Feed Forward 15*	Lime Slurry Pumped 17	Reactor Lime Feed 18*	Reactor Effluent 19*	Filter Feed 20	Thickener Overflow 21	Cake Wash Water 22
H_2O, #/min	13,642.8	14,828.1	2,150.8	537.7	14,233.0	2,103.5	29,182.8	756.0
Na_2CO_3, #/min		6.8					13.9	
NaOH, #/min		50.0			50.2	7.3	101.3	
Na_2SO_3, #/min	119.7	347.2			349.4	51.0	704.9	
$NaHSO_3$, #/min	456.4							
Na_2SO_4, #/min	1,056.0	1,016.3			1,019.8	143.7	2,057.6	
$Ca(OH)_2$, #/min			852.0	213.0				
$CaSO_3 \cdot 1/2\ H_2O$, #/min					331.0	662.0		
$CaSO_4 \cdot 1/2\ H_2O$, #/min					39.4	78.8		
$CaCO_3$, #/min					5.8	11.6		
Inerts, #/min			69.2	17.3	17.3	34.6		
NaCl, #/min	305.1	303.6			305.1	44.5	615.5	
Total, #/min	15,580.0	16,552.0	3,072.0	768.0	16,351.0	3,147.0	32,676.0	756.0
% Solids	0	0	30.0	30.0	2.4	25.0	0	0
Flow, GPM	1,698.0	1,837.0	305.2	76.3	1,785.0	302.0	3,618.0	91.0
pH			10-14	7.0	6-13	7-13	7-13	
Temp, °F					120-140	120-140	120-140	AMB

Stream No.	Cake 23	Filtrate 24	Soda Ash Solution 25	Soda Ash Feed 26	Soda Ash Solution Liquor 27	Make-up Water 28	Lime Slurry Recycle 29	Underflow Slurry Dilution Liquor 30	River Water (Startup) 33
H_2O, #/min	442.1	2,421.4	403.9		403.9	878.0	1,075.4		
Na_2CO_3, #/min			13.9	13.7	0.2				
NaOH, #/min	0.50	6.8	1.4		1.4				
Na_2SO_3, #/min	3.4	47.5	9.6		9.6				
$NaHSO_3$, #/min									
Na_2SO_4, #/min	10.0	138.8	27.9		27.9				
$Ca(OH)_2$, #/min							426.0		
$CaSO_3 \cdot 1/2\ H_2O$, #/min	662.0								
$CaSO_4 \cdot 1/2\ H_2O$, #/min	78.8								
$CaCO_3$, #/min	11.6								
Inerts, #/min	34.6						34.6		
NaCl, #/min	3.00	41.5	8.3		8.3				
Total, #/min	1,246.0	2,656.0	465.0	13.7	451.3	878.0	1,536.0		
% Solids	63.1	0	0				30.0		
Flow, GPM		299.0	50.0		50.0	105.0**	152.6	250.0***	135.0 max
pH		7-13	8-14		7-13			7-13	
Temp °F		120-140	120-140		120-140	AMB		120-140	

* Per module

** Includes water entering process from pump seals

*** Intermittent Maximum

Combustion Equipment Assoc.
Material Balance
for
Drawing 040044-1-2 Rev G

inlet dampers have been interlocked to enable bringing the absorbers on - or off-line without interruption of the boiler operation.

Hot flue gas (300°F) entering each absorber is first cooled by a set of sprays which direct scrubbing solution to the underside of the bottom tray. In addition to providing temperature control at the bottom of the absorbers, these sprays are intended to keep the underside of the tray and the bottom of the absorber free of fly ash solids buildup. The cooled, saturated gas passes through a set of two trays where SO_2 is removed and then through a chevron-type mist-eliminator, to remove any entrained absorber liquor droplets. After leaving the absorber, the scrubbed gas is reheated 50F° (to a temperature of about 175°F) by mixing with hot air from an oil-fired reheater to prevent condensation and corrosion.

After the scrubbing solution passes the internal contacting area (two single-pass crossflow perforated plates), it is collected at the bottom of the absorber. Part of the liquor is recirculated to the spray and to the top of the trays. The recirculating liquor and regenerated liquor are flow-mixed prior to entering the top of the tray.

By adjusting the flow rate of the feed forward, the tray pH is controlled below 10 pH to minimize the absorption of CO_2 and decrease the potential for $CaCO_3$ scaling in the absorber.

Provisions have been made for an automatic emergency water supply to spray and quench the hot flue gas in the event of recycle pump failure. Pressures and temperatures are measured at appropriate points throughout

the system. Removal of SO_2 is monitored by continuous SO_2 and O_2 analyzers at the inlet of the absorbers and in the stack. The pH of the spent liquor from the absorber and of the feed to the top tray are also continuously monitored.

Reactor Section - This section of the system consists of two parallel reactor trains each containing a primary and secondary back-mix reactor. At operation under design conditions (i.e., full load) each reactor train handles the regeneration of solution from one absorber. For short periods of time, a single reactor train is capable of handling the solutions from both absorbers operating at full load. When the boiler is firing typical or average coal (3.5% sulfur), only one reactor train is normally required.

The spent scrubbing solution is introduced to the primary reactor along with slurried carbide lime from the lime day-tank. In this reactor (3-15 minutes residence time), the regeneration of scrubbing solution begins. The primary reactor overflows into a larger reactor (30-60 minutes residence time) where regeneration is completed. The secondary reactors are designed to maintain a pH in the range of 11 to 12 by controlling the amount of lime slurry fed to the primary reactors.

Solids Dewatering - The slurry from the secondary reactor is fed to the thickener feed well where separation of the regenerated solution is initiated. The thickener is operated to provide 15-30 wt % solids, even though the slurry can settle to 40 wt % solids or more. Slurries of about 25 wt % solids allow better control of filter cake washing, control

which cannot always be achieved with variations in filter drum speed and pool depth. The thickener underflow slurry is recirculated past the filters in a recycle loop that feeds a portion to the filters and returns the remainder to the thickener feed well. Each filter is equipped with an overflow pipe returning to the solids zone in the thickener to allow for operation in an overflow mode.

There are three filters, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (i.e., to obtain cake containing high percetage dry solids and low percentage soluble salts) the filters should run at fixed conditions (constant drum speed, submergence, wash ratio, etc.). Therefore, the cake discharge rate is controlled by changing the number of filters in operation. The number of filters to be placed on line is judged from the quantity of solids accumulated in the thickener. This in turn is judged from the solids concentration in the underflow slurry and the thickener rake lift position indicator. The density of the underflow slurry is measured and thickener hold tank liquor is added at the thickener outlet as required to maintain the solids concentration in the underflow slurry at about 20-25%. The number of filters in operation is changed if the concentration of solids in the underflow slurry cannot be controlled using the dilution liquor.

The solid cake is washed on the filter using a series of water spray banks. This wash is designed to remove a large fraction (approximately

90%) of the occluded soluble salts from the cake and to return these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup. The total wash rate will usually be set at a constant percentage of the cake discharge rate. The mixed filtrate and wash liquor from the filter is collected in the filtrate sump and returned to the thickener.

Clear liquor overflow from the thickener is collected in the thickener hold tank which provides surge capacity for the absorbent liquor feed to the scrubber system and maintains overall control of the volume of liquor in the system. Water is added to this hold tank to make up for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (pump seals, lime feed, and cake wash).

Raw Materials Preparation - Two chemicals are required for the operation of the dual alkali system at Cane Run Unit 6, carbide lime for liquor regeneration and soda ash to make up of sodium losses.

Carbide lime, a byproduct of acetylene production is barged to the Cane Run Station as a slurry containing approximately 25% insoluble solids. The slurry is hydraulically unloaded and pumped to a main storage tank at the plant. Excess liquor is then decanted from the tank and returned for reuse in transport of carbide lime. From the main storage tank, the lime slurry is pumped to a grinding system consisting of a hydrocylone and wet ball mill (to prevent feeding of oversized material). From the grinding system, the lime is pumped to the dual al-

kali system day-tank which supplies lime to the primary reactor at the appropriate rate.

The system was designed to utilize 95% of the available $Ca(OH)_2$ in the carbide lime slurry. Thus the overall feed stoichiometry (moles available $Ca(OH)_2$ per mole of SO_2 absorbed) is expected to be 1.05. The specifications for the carbide lime as delivered to the dual alkali system are given in Table 2.3.

Dry, dense soda ash is received at the plant and stored in the soda ash silo which feeds the soda ash solution tank by means of a weigh feeder. The rate of addition is proportional to the amount of lime slurry fed to the system. Such relationship arises from the fact that the sodium losses in the cake can be expressed as a percentage of the solids produced, which in turn is related to the amount of lime added to the system.

Soda ash solution can be made up using either clarified liquor drawn from the thickener hold tank or fresh water. Provisions have also been made to feed the soda ash solution either directly to the absorbers or to the thickener. The normal mode of operation is to prepare the soda ash make-up solution using clarified liquor and to feed it to the absorbers.

Waste Disposal- A long-range plan for disposal of the dual alkali filter cake has been developed as part of an overall disposal plan for all FGD wastes produced at Cane Run. The plan involves stabilization of the wastes generated by each of the three FGD systems via the addition of lime and fly ash and then dry landfill of the stabilized material. Waste from the dual alkali system on Unit 6 will be handled, processed,

Table 2.3

Carbide Lime Specifications[a]

	Weight Percent Based on Dry Solids
$Ca(OH)_2$ (equivalent to 70.01% CaO)	92.50
$CaCO_3$	1.85
SiO_2	1.50
$(Fe_2O_3 + Al_2O_3)$	1.60
MgO	0.07
S	0.15
P	0.01
Free carbon	0.25
Not analyzed	2.07

[a]Available as slurry containing 30% insoluble solids.

Source: Airco catalog (1969)

and disposed of independently of the wastes produced by the direct lime scrubbing systems on Units 4 and 5. Wastes from these latter two units will be combined and handled in a common processing plant.

The waste processing plant, IU Conversion System (IUCS), is installed at the end of the filter building. Filter cake from the dual alkali system discharges into a single conveyor (the three filters are arranged in a single-row end-to-end), which carries the cake directly to the pug mill mixer in the processing plant. A belt weigh element near the end of the conveyor measures the quantity of cake produced. Fly ash from Unit 6 is pneumatically transported to the processing plant and stored in a live-bottom silo, from which it is discharged to a screw conveyor and fed to the pug mill. The fly ash is adjusted according to the filter cake rate.

Quicklime is also fed to the pug mill at a rate equivalent to a few percent of the combined dry weight of ash and cake. The lime is stored in a dry silo above the pug mill.

The pug mill discharges to a radial stacking conveyor. The stacker spreads the material in a windrow (interim stockpiling area) where it remains for 1-3 days to initiate the stabilization reactions. The material is then removed from the windrow and loaded into rear dump trucks using a frontend loader. The trucks transport the material to a dry landfill south of the plant on LG&E property. Provisions have also been made to allow direct loading of the trucks from the stacking conveyor, should this mode of operation prove more satisfactory.

2.4 Design Operating Conditions for LG&E Dual Alkali System

The operating conditions for which the LG&E system was designed are presented in Table 2.4. As will be explained in more detail in Section 3 of this report, the system was not operated under these conditions.

The presence of soluble aluminium compounds in the carbide lime (and fly ash carried over to the absorber) has meant that the pH's in the secondary reactors must be limited to below 10 and in the thickener to below 10.5.

2.5 Final LG&E Dual Alkali System Design and Cost

A complete description of final design of the Louisville Gas and Electric balances, equipment specifications, as well as the final system cost appears in the Phase II report to EPA, Reference (1). Tables 2.5 through 2.7 are abstracts from this report.

Table 2.4

Design Operating Conditions* For LG&E Dual Alkali System

Absorber:	Flow rate to absorber 280-1850 gpm Spent liquor pH: 6.2 Top Tray Feed pH: 7.5-10.5
Primary Reactors:	Flow rate: 260-1700 gpm Retention time: 3-15 min. pH: 5-11
Secondary Reactors:	Flow rate: 260-1700 gpm Retention Time: 30-60 min pH: 11-12
Thickener:	Flow rate: 600-3600 gpm Retention Time: 8.5-51 hrs pH: 12
Filters:	Drum Speed: 0.5-0.7 rpm Wash Rate: 40-50 gpm Cake thickness 1/4 - 3/8 inches

*Reference (1)

Table 2.5

Design Basis*

Coal (Dry Basis):	
Sulfur (% S)	5.0
Chloride (% Cl)	0.04
Heat Content (Btu/lb)	11,000
Inlet Gas:	
Flow Rate (acfm)	1,065,000
(lb/hr)	3,372,000
Temperature (°F)	300
SO_2 (ppm-dry basis)	3,471 ppm
O_2 (vol. %)	5.7
Particulate Matter (lb/MM Btu)	<0.10
Outlet Gas:	
SO_2 (ppm - dry basis)	<200
(lb/MM Btu)	0.45
Particulate Matter (lb/10^6 Btu)	<0.10

* Reference (1)

Table 2.6

Basis for Material Balances at Design Conditions*

Coal (dry basis):	
Sulfur (% S)	5.0
Chloride (% Cl)	0.04
Heating value (Btu/lb)	11,000
Sulfur volatilized (% S in Coal)	94
Inlet Gas:	
Flow rate (acfm)	1,065,000
(lb/hr)	3,372,000
Temperature (°F)	300°F
SO_2 (ppm-dry basis)	3,471
O_2 vol. %	5.7
Particulate Matter lb/10^6 Btu	<0.1
Outlet Gas:	
SO_2 (ppm dry basis)	200
(lb/MM Btu)	0.45
Particulate Matter lb/10^6 Btu	<0.10
Absorber Feed Concentration (Molarity):	
Na^+ associated with OH^- and $CO_3^=$	0.09
$SO_3^=$	0.36
Oxidation and Sodium Makeup Rates:	
Oxidation (molar basis)	10%
Na_2CO_3 moles Na^+/mole SO_2	0.045
Calcium Feed:	
Solids in slurry	30%
Available $Ca(OH)_2$	92.5%
$Ca(OH)_2$ utilization	98%
Waste Solids:	
Wash ratio	2/1
Insoluble solids	63%

*Reference (1)

Table 2.7

Capital Costs* for the Dual Alkali System
at Cane Run Unit 6, LG&E
"as-incurred Plus Estimate for Completion Basis"

	(x1000 Dollars)
MATERIAL COSTS ($):	
o FGD System	
- Process Materials	7,392.3
- Additional Materials	
. Utility and Service Piping and Instrumentation	228.9
. Electrical	731.7
. Other	1,740.3
	10,093.2
o Lime Slurry Feed System	
- Process Equipment	617.2
- Electrical, Piping and Instrumentation	170.8
	788.0
o Solid Waste Disposal System[a]	
- Fly Ash Supply	309.6
- Waste Processing	1,060.9
- Landfill Area Materials[b]	559.0
	1,929.5
TOTAL MATERIAL COSTS	12,810.7
ERECTION COSTS ($):	
o Direct Labor	3,058.5
o Field Supervision[c]	337.2
o Construction Overhead	2,038.9
TOTAL ERECTION COSTS	5,434.6
ENGINEERING COSTS ($):[c]	
o System Supplier's Engineering	1,162.7
o Owner's Consultant Engineering	985.0
TOTAL ERECTION COSTS	2,147.7
SPARE PARTS:	203.9
TOTAL CAPITAL INVESTMENT	20,596.9

a Battery limits: starting at cake discharge from filters.
b Includes earth moving equipment, dikes, culverts, and bridges.
c Owner's engineering is included in field supervision.

* Reference (1)

Section 3

OPERATING HISTORY OF LG&E DUAL ALKALI SYSTEM FROM INITIAL START-UP TO ACCEPTANCE TEST

3.1 Start-up and Initial Operation -- March 1979 through April 1980

Figure 1.1, page 4, presented the original and the actual schedule of the Acceptance Testing and yearlong Demonstration Program (Phase III & IV).

3.1.1 Chronology

Table 3.1 summarizes the operating history of the dual alkali system at LG&E from the initial filling with scrubbing liquor (end of Phase II), through Acceptance Testing (end of Phase III). Also included are the first three months of the Demonstration Year (completed prior to the Acceptance Testing during a period of reduced boiler load which had delayed the start of Acceptance Testing).

3.1.2 Problems and Solutions

A number of problems were encountered during the start-up and initial operation of the system. This section summarizes the most significant of these and discusses the steps taken to solve them.

Soda Ash Solution Makeup

In normal operation, makeup soda ash is dissolved in thickener return water (which has a temperature in excess of 100°F) and then returned

Table 3.1

Summary of Operating History of LG&E Dual Alkali System
From Initial Fill to Acceptance Test

Month	Boiler Hours	Dual Alkali Scrubber Hours	Comments on Dual Alkali System
3/79	---	---	Start fill of dual alkali system
04/79	---	---	Start-up of absorbers
05/79	578	164	Low-lime solids, leak in absorber shell
06/79	602	74	Mist eliminator collapse
07/79	528	0	Awaiting delivery of new mist eliminators
08/79	693	51	Installation of new mist eliminators, low lime solids
09/79	530	300	Pump impeller replacement, repair of bypass damper
10/79	247	217	Level tray overflow weirs, tray scaling
11/79	468	456	Scrubber cleaned, SO_2 monitors calibrated
12/79	701	667	First attempt to conduct Acceptance Test
01/80	210	104	Thickener pluggage and jammed thickener rake
02/80	386	327	Thickener (and absorber) cleaning
03/80	686	481	Carbide lime shortage, ESP testing, IUCS tie-in
04/80	581	531	Repairs to IUCS system, thickener, underflow line
DUAL ALKALI SYSTEM LINE-OUT PERIOD PRIOR TO ACCEPTANCE -- May-July 1980 (Phase IV-Demonstration Year Testing)			
05/80	741	740	Boiler load restricted by feed water pump
06/80	642	640	Testing of alkali consumption and calibration of SO_2 monitors
07/80	518	517	Final preparation for Acceptance Test and Acceptance Test

to the scrubber. During the initial filling of the dual alkali system in March 1971, however, the soda ash did not dissolve easily because of the low temperature (50 - 54°F) of the fresh makeup water supplied to the soda ash tank. In order to alleviate this problem, a small package boiler was used to preheat the water before it was pumped to the soda ash tank.

Low Carbide Lime Solids Concentration

Supplying carbide lime with a solids concentration above 20% to the dual alkali system was a problem until modifications were made in the carbide lime barge loading system at Airco (the carbide lime supplier) and the lime grinding/classifying system at LG&E in February and March 1980. Excess water introduced with the lime at low solids concentration upset the water balance of the scrubbing system, since the water input with lime feed and filter cake wash exceeded the water losses in the filter cake and scrubber evaporation. The resultant increase in water inventory eventually filled the thickener to the overflow level. At this point, it was necessary to discontinue the lime feed until enough water had been evaporated from the scrubbing liquor by the hot flue gases in the scrubber to drop the thickener level to normal. Discontinuation of lime feed caused the process scrubbing liquor to become acidic and SO_2 removal efficiency dropped below the required 90%.

A temporary solution involved decanting the dilute lime slurry (from the barges) in the lime storage tanks, and pumping the set-

tled slurry to the day-tank. Because lime requirements for the scrubbers could not always be coordinated with barge unloading schedules, high solids lime could not always be supplied to Unit 6 and there were still periodic problems with water balance.

Since Airco completed its larger pipeline for delivery of concentrated lime to the transport barge and the LG&E carbide lime grinding and classifying system became fully operational, there has not been a problem in maintaining an adequate concentration of carbide lime solids in the feed to the dual alkali system at LG&E.

Chevron Mist Eliminators

After several weeks of initial scrubber operation, an increase in pressure drop across both the scrubber modules was noted. The scrubbers were inspected and it was discovered that in addition to some deposit formation on the trays, the chevron type Noryl® mist eliminator had deformed closing the gas passage. An attempt was made to repair repair the mist eliminators on site, but this approach was finally abandoned and a new set of chevron mist eliminators were ordered from a different supplier, Ceilcote. These were also fabricated from General Electric's Noryl® plastic, but the new chevrons had considerably more support bracing and no further difficulties have been experienced with the mist eliminators.

Pump Failure

Two major changes in the Worthington scrubber recycle and thickener overflow return pumps have been made since scrubber startup. These

were: (1) replacement of the two-piece welded impellers with single-piece cast impellers and (2) replacement of the two-piece suction-side pump liner with a bonded liner.

The welded type impeller was replaced because of a high incidence of weld failure (shearing between the hub and impeller) and the two-piece pump liner had allowed leakage to corrode the bolts holding the liner in place, resulting in the liner being drawn into the impeller.

By-Pass Damper Leakage

Gas leakage through the scrubber by-pass damper has consistently exceeded specification (1%). This has periodically resulted in recirculation of scrubbed gas through the system and by-pass of unscrubbed flue gas to the stack. Linkages to the damper louvers have been adjusted a number of times and an additional set of actuators and limit switches was installed in January 1980. The additional actuators and limit switches have helped reduce by-pass damper leakage, but not down to the 1% leakage specified.

Tray Overflow Weirs

During a routine inspection of the scrubbers while the boiler was being repaired, it was discovered that the liquid distribution weirs in both absorbers were not level. Since this could lead to uneven liquor distribution, poor SO_2 removal efficiency and dry areas on the tray, the weirs were releveled and stainless steel strapping was welded across the weirs to hold them in position.

Tray Scaling

Formation of scale deposits on the absorber trays had been evident from the time of startup of the scrubbing system. Whenever an outage of the boiler or scrubbing system permitted, the trays were recleaned. When scrubber operation resumed, so did scale formation. Eventually enough of the holes in the trays would become blocked and pressure drop across the scrubber would become high enough to require a shutdown for cleaning of the trays.

Arthur D. Little and Bechtel analyses of the tray deposits showed that the scale was mainly a complex of alumina and silica. Subsequent testing by Arthur D. Little demonstrated that alumina in the carbide lime dissolved at a pH of 11.4 and precipitated when the pH dropped below 10.7.

The original design and operating conditions for the system called for pH of:

- 6.0 for the absorber,
- 11.5 for the secondary reactor,
- 12 for the thickener overflow, and
- 7 - 10.5 for combined thickener overflow and absorber recycle stream.

These conditions were ideal for dissolving aluminium in the reactors and thickener and then precipitating the alumina-silica complex on the absorber trays.

As a result of these findings, the operating pH conditions were changed

from those originally specified in the design (see Table 2.4) to those summarized below:

- 6.3 for the absorber,
- 8.0 for the primary reactor,
- 9.0 for the secondary reactor, and
- <10.5 for the thickener overflow.

To compensate for the lower pH of the thickener overflow being fed to the top tray (and to maintain SO_2 removal efficiency), it was necessary to increase the pH of scrubber liquor being recirculated from the bottom of the scrubber module to the trays. This, in turn, was accomplished by increasing the operating pH of the absorber module by increasing the "feed forward" flow of liquor from the thickener to the absorber.

SO_2 Removal Efficiency Monitoring

Shortly after startup of the scrubbing system, a difference between the SO_2 removal efficiencies of Modules A and B (as measured by the DuPont monitors in the scrubber inlet and outlet) was noted despite the fact that both absorbers were operating under the same conditions of flow and pH. This difference in SO_2 removal efficiency was not affected by recalibration of the monitors. Switching the connections between the SO_2 probes and DuPont analyzers of Modules A and B indicated it was the outlet SO_2 probe location, rather than the monitors, that was the problem. Subsequent testing of gas velocity and SO_2 concentration profiles across the scrubber outlet ducts by LG&E con-

firmed that stratification of gas flow in the outlet duct was causing the differing SO_2 concentrations across the outlet duct. The SO_2 probes were, consequently, relocated downstream in the scrubber outlet ducts.

Fly Ash Carryover

The Unit 6 dual alkali scrubbing system was designed for an inlet (fly ash) particulate loading of <0.1 lb/10^6 Btu. Fly ash particulate loading to the scrubber has averaged about 0.85 lb/10^6 Btu, over eight times the design figure. During upsets in electrostatic precipitator (ESP) operation the loading was often significantly above this average.

Although the scrubbers were not designed for particulate removal, they were very effective in removing a high percentage of the fly ash from the flue gas. Since the scrubber modules are designed for clear liquor operation, there are no agitators in the modules and fly ash solids accumulated in the bottom of the scrubber. In January 1980, during a shutdown of the scrubbing system for work on the by-pass damper, both absorber modules were cleaned of accumulated fly ash solids and the solids flushed to the thickener. The thickener rake jammed due to the overload of solids and it was necessary to drain the thickener. A hole was cut in the side of the thickener and the solids removed using a front-end loader.

In addition to the solids accumulation problem, the fly ash was also found to contribute dissolved aluminium (possibly with the potential for scaling) at a pH of over 11.0.

While excessive fly ash carryover from the ESP during the malfunction

caused pluggage of the scrubber lines and valves and increased the load on the thickener, the scrubbing system did seem able to tolerate the average Unit 6 ESP outlet fly ash loading. After a stabilized bed of fly ash solids formed on the scrubber module floor, excess fly ash carried over to the thickener.

Scrubbing Liquor Regeneration Reactors

When lime feed to the regeneration reactors was initiated, the current drawn by the agitators in both the primary and secondary reactors exceeded safe limits for the motor drives. Examination of the design criteria revealed that the motor and drives were undersized for the agitator design. Since replacement of both motors and drives would have been costly and would have delayed startup, the impellers were modified to reduce power consumption.

It was then found that because of the direction of rotation of the agitator and the location of the overflow pipe in the secondary reactors, most of the flow from the primary reactors was passing directly to the overflow of the secondary reactor. This short circuiting was eliminated by the installation of a 16-foot chute (down-pipe) fastened to the reactor baffles. This chute delivered input flow from the primary reactor to the area near the secondary reactor impeller thereby providing much more efficient mixing within the tank.

Originally, the secondary reactors were designed to operate at different liquid levels to allow adjustment of retention time. A combination of level controllers, discharge pumps, and discharge valves

were to be used to control reactor liquid level. However, the pressure developed by the discharge pumps was more than the discharge valve could handle without violent vibration at low flow. The secondary reactors were, therefore, operated on overflow rather than level control.

By operating the secondary reactors on overflow to the thickener, maximum retention time was obtained. (It was subsequently found that this maximum retention time was required for completion of the reaction between the carbide lime and scrubbing liquor.)

Soda Ash Feeder

Several problems were encountered with the dry soda ash feed. Moisture entering the silo, particularly in damp rainy weather, formed hard lumps or "rocks" of soda ash that jammed the feeder. Numerous attempts were made to seal the silo against moisture, but there was still some leakage and occasional jamming of the feeder with hard lumps of soda ash.

A small blower was installed in the chute between the feeder and the dissolving tank, preventing moisture (from the dissolving tank) from caking the soda ash on the side of the chute and blocking it.

A problem with the zeroing circuit of the weighing mechanism of the soda ash feeder was also finally corrected.

Fiberglass Piping

There were a number of serious failures of the fiberglass piping and

fittings. Most of these breaks occurred in areas where there was vibration or mechanical stress and include breaks in the absorber recycle lines, thickener return and underflow lines. In some critical areas, such as the thickener underflow lines in the tunnel (where a break flooded the tunnel with three feet of slurry) the fiberglass piping was replaced with carbon steel.

3.2 First Attempt to Conduct Acceptance Test

On December 3, 1979, the first attempt was made to conduct the Acceptance Test. On December 7, 1979, the official Acceptance Test was terminated by LG&E because of:

- low solids concentration in the carbide lime feed,
- high pressure drop across the absorber, and
- lack of (operable) spare recycle pumps.

Although the official testing had terminated, unofficial testing continued until December 11 to gather operating data. At that time, the system was shut down for lack of carbide lime feed.

3.3 Line-Out Period and Start of Demonstration Year, May-July 1980

Under the original agreement and schedule, it was intended that the Acceptance Test be carried out in advance of the Demonstration Year. In late April, however, after repeated postponement of the Acceptance Test, it was decided to proceed with the Demonstration Year beginning May 1, 1980 despite the delay in Acceptance Testing. The Acceptance Test was to be rescheduled for a later time when all Unit 6 and FGD

system components could be fully and simultaneously available and the boiler operated at, or near, full load.

Boiler load was initially limited to 180 Mw in May due to outage of a boiler feed water pump. Repair of the pump at the beginning of June allowed the boiler to reach 240 Mw, the capacity at which it was agreed to conduct the Acceptance Test in July.

However, several problems which could have resulted in an unsuccessful Acceptance Test were also uncovered in June. Bechtel determined that alkali consumption, both sodium and calcium, was more than 30% over the level stated in the guarantees. In addition, the difference in apparent SO_2 removal efficiency between absorbers A and B raised the question of SO_2 analyzer accuracy.

To measure the effect of wash rate on sodium consumption, filter tests were conducted by LG&E, Bechtel, and Arthur D. Little. Sodium analyses of the cake indicated that the efficiency of filter cake washing corresponded to dilution of the sodium liquor in the cake rather than displacement of the liquor. Therefore, the wash rate was increased to about 65 gpm instead of the 40-50 gpm specified originally.

The cause for the discrepancy in the absorber outlet SO_2 readings, namely non-uniform gas velocity in the scrubber outlet ducts, was also determined before the Acceptance Test. To give a more representative SO_2 reading, the absorber outlet SO_2 DuPont probes were repositioned in the absorber outlet ducts. It was also agreed that Lear-Siegler analyzer in the stack would be the primary analyzer for the Acceptance Test and outlet DuPont analyzers would be used only for backup.

The performance factors for the first three months of the Demonstration Year (May, June, July) are given below:

	May	June	July
Availability	99.9%	99.9%	99.9%
Operability	99.9	99.7	99.8
Reliability	99.9	99.9	99.9
Utilization	99.5	88.9	82.9

Section 4

ACCEPTANCE TEST

4.1 Performance Guarantees and Test Results

The Acceptance Test was conducted to assess the performance of the dual alkali system with respect to guarantees offered LG&E by CEA/ADL and passed through in the contract between EPA and LG&E. There are seven guarantees applying to: SO_2 emissions, particulate matter emissions, lime consumption, soda ash consumption, power consumption, waste solids properties, and system availability. Table 4.1 summarizes those guarantees and corresponding results of the Acceptance Test. A brief discussion of each guarantee test performed during the Acceptance Test follows. The actual guarantee language from the CEA-LG&E contract is given in quotations.

4.1.1 Sulfur Dioxide Emissions

"The [FGD] system shall provide such control that emissions from the stack shall be no greater than 200 ppm by volume SO_2 dry basis (D.B.) not including SO_2 added from the operation of the reheaters and without additional air dilution when burning coal containing less than 5% sulfur. When burning coal containing 5% sulfur or greater the [FGD] system shall be designed to remove at least 95% of the sulfur dioxide in the inlet flue gas."

Normal air dilution was based on historical oxygen data and on oxygen measurements taken during the acceptance test. The correction for

Table 4.1

Performance Guarantees and Acceptance Test Results

GUARANTEE	TEST RESULTS
SO_2 Removal	
200 ppm dry basis (D.B.) without additional air dilution	143 ppm (D.B.) without additional air dilution
Calcium Consumption	
1.05 moles available CaO per mole SO_2 removed	1.04 moles available CaO per mole SO_2 removed
Soda Ash Consumption	
0.045 moles Na_2CO_3 per mole SO_2 removed	0.042 moles Na_2CO_3 per mole SO_2 removed
Net Particulate Addition	
No net particulate addition by FGD system	Net particulate removal averaging 88% efficiency
Power Consumption	
System will consume (excluding reheat) not more than 1.2% of power generated at peak capacity	System consumed 1.05% of power generated
Filter Cake Properties	
Filter cake will contain a minimum of 55 wt % insoluble solids	Filter cake averaged 52.2 wt % insoluble solids

additional air dilution" was made using the following formula:

SO_2 (Corrected) = SO_2(Stack) x [20.9 - O_2 (Inlet)]/[20.9 - O_2 (Outlet)]

Where:

SO_2 (Corrected)	=	Actual outlet SO_2 concentration measured at the stack and corrected for dilution by reheater exhaust, ppm
SO_2 (Stack)	=	SO_2 outlet concentration corrected to dry, ppm
O_2 (Inlet)	=	Percent oxygen concentration at inlet to scrubber
O_2 (Outlet)	=	Percent oxygen concentration in the stack

The primary method of determining SO_2 removal relied on the continuous Lear Siegler monitor (SM810) installed in the stack and the DuPont Model 460 analyzer installed at the inlet. The stack analyzer was certified in December 1979 by an outside contractor according to the procedure specified in the Federal Register. During the Acceptance Test, as a backup to the continuous monitors and as an ongoing confirmation of analyzer accuracy, wet chemical tests of the stack effluent and system influent were performed daily, according to EPA Method 6 in conjunction with the particulate matter tests.

Preliminary results from the wet chemical analysis showed a discrepancy between these wet chemical analyses and the Lear Siegler monitor read-out. After an extensive check of the system, a burned ground wire was discovered in the signal line of the Lear Siegler continuous SO_2 monitor. From the data on the calibration sequences of the analyzer prior to, during, and after elimination of the grounding problem, it was concluded that the analyzer signals were indicating 30 ppm lower than

actual ppm. Therefore the continuous stack SO_2 monitor readings for the first 7 days of the tests were corrected by 30 ppm. With this correction applied to the early readings, and subsequent to the repair to the ground in the analyzer, the two techniques were in good agreement. Table 4.2 summarizes the 24-hour Average SO_2 results for the Acceptance Test. Table 4.3 summarizes the simultaneous wet chemical and continuous monitor measurements (the Method 6 tests were conducted only for the first 10 days). Appendix A contains detailed results of the Acceptance Test analyses on a daily basis to which the reader is referred for further information.

4.1.2 Particulate Matter Emissions

"In addition to meeting applicable regulations, the [FGD] system will also be designed to meet Federal New Source Performance Standards for emissions of particulate matter under all conditions of boiler operation. The dual alkali system will not add any particulate matter to the emissions of particulate matter that are received by the system from the LG&E Cane Run No. 6 electrostatic precipitator."

The Federal New Source Performance Standards for particulate matter emissions are: 1) 0.1 lb particulate matter per million Btu heat input is the maximum two hour average allowed, and 2) the opacity shall not exceed 20% (excluding water), except that 40% opacity is permissible for not more than 2 minutes in any hour [Reference (2)].

Addition of particulate matter by the D/A system is interpreted to mean that the total particulate mass loading after the D/A system shall not

Table 4.2

Acceptance Test Continuous SO_2 Analysis

Acceptance Test Day	24 Hour Continuous SO_2 Analyzer Results (ppm, dry basis)				L/G	Act. Na
	A Inlet	B Inlet	Stack	% Removal	$\frac{gal}{10^3 acf}$	(M)
1	2444	2418	130	94.7	9.5	0.42
2	2674	2570	129	95.0	8.7	0.40
3	*	2390	130	94.6	9.4	0.42
4	*	2290	152	93.4	9.2	0.44
5	2265	2315	157	93.1	9.1	0.54
6	2567	2515	140	94.5	7.8	0.52
7	2113	2021	132	93.6	8.3	0.49
8	2116	2088	124	94.1	8.9	0.49
9	2395	2339	146	93.8	8.2	0.49
10	2372	2315	171	92.7	8.7	0.49
11	2292	2233	156	93.1	9.6	0.43
12	2167	2166	130	94.0	9.8	0.43
Average	2340	2305	141	93.9	8.9	0.46

* Analyzer printout malfunction

Table 4.3

Acceptance Test Continuous Monitor and EPA Method 6 Analysis

Acceptance Test Day	Hours	SO_2 Concentration, ppm, dry basis					
		A Inlet		B Inlet		Stack	
		DuPont Analyzer	Method 6	DuPont Analyzer	Method 6	Lear Siegler Analyzer*	Method 6
1	1400-1700	2434	2330	2516	2330	119	124
2	1100-1300	2434	2150	2423	2180	122	163
3	1000-1300	2592	2210	2670	2290	117	154
4	1200-1500	2836	2390	2674	2480	155	159
5	1000-1300	2656	2330	2606	2410	136	137
6	1000-1300	2716	2350	2418	2480	184	212
7	1600-1900	2337	2040	2250	2030	**	137
8	1100-1300	2395	2120	2330	2100	113	130
9	1100-1400	2864	2530	2721	2450	122	133
10	0900-1200	2690	2410	2624	2360	160	137

* Lear Siegler analyzer readings for days 1-6 corrected for the effect of the burned out ground wire

** Analyzer out of service for repairs to ground wire

be greater than the mass loading before the D/A system. Particulate tests, following EPA Method 5, were conducted on the inlets to the absorber modules and in the stack (downstream of reheaters) during the Acceptance Test. The results of 10 simultaneous tests showed that there was no net addition of particulate matter across the system. Actually, the absorber performed as particle removals devices averaging 88% net removal of incoming particulate. Table 4.4 displays the results of particulate matter tests performed during the test program.

Although the FGD system met the guarantee requirements, the test was not very stringent due to the low level of performance by the electrostatic precipitator during the acceptance test period. The FGD system was originally designed for an incoming flue gas stream containing the equivalent of 0.1 lb of particulate matter/10^6 Btu or less. During the Acceptance Test, however, the level of incoming particulate matter was almost an order of magnitude higher. Thus it is not surprising that the absorbers functioned to remove particulate matter even at the relatively low pressure drop at which they operated. The particulate matter emissions from the stack, however, were on the order of 0.1 lb/10^6 Btu as required for Unit 6 FGD system under the appropriate requirements to control particulate matter emissions.

4.1.3 Lime Consumption

"The consumption of lime in the [FGD] system will not exceed 1.05 moles of available CaO in the lime feed per mole of SO_2 removed from the flue gas". The lime consumption guarantee may be exceeded by 30% during Acceptance but must be met during the Demonstration year.

Table 4.4

Acceptance Test Particulate Test Results

Acceptance Test Day	Particulate ($lb/10^6$ Btu)			
	A Inlet	B Inlet	Stack	% Removal
1	0.5320	0.7120	0.0895	85.6
2	0.6590	0.3620	0.0932	81.7
3	0.9470	1.0700	0.1110	89.0
4	0.9440	0.8060	0.1030	88.2
5	1.1100	0.9200	0.1020	90.0
6	0.9900	1.4900	0.1020	91.8
7	0.5890	0.8470	0.1020	85.6
8	0.6250	0.6490	0.0893	86.0
9	0.7890	1.2000	0.1100	88.9
10	0.9620	0.5930	0.1020	86.9
Average	0.8147	0.8649	0.1004	88.0

During the 12-day Acceptance Test the calcium consumption, corrected for available alkalinity averaged 1.04 moles of available CaO per mole of SO_2 removed, thus meeting the guarantee which required less than 1.05 moles per mole of sulfur removed.

Table 4.5 summarizes the analyses performed on the filter cake samples. Appendix A presents a summary of analytical results from all filter cake samples taken during the Acceptance Test.

4.1.4 Sodium Carbonate Makeup Guarantee

"Soda ash makeup will not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas provided that the chloride content of the coal burned averages 0.06% or less. If the average chloride content of coal is above 0.06%, then additional sodium carbonate consumption will be allowed at the rate of 1/2 mole Na_2CO_3 for each mole of chloride in the flue gas resulting from chloride in excess of 0.06% in the coal. The soda ash consumption guarantee may be exceeded by 30% during Acceptance Testing, but must be met during the Demonstration year".

The soda ash makeup was determined by analysis of total sodium and total sulfur in the filter cake. According to this analysis the soda ash make-up averaged 0.042 moles of Na_2CO_3 per mole of sulfur dioxide removed and therefore met the guarantee requirement of 0.045. Analysis of the as-fired coal showed the chloride concentration to be below 0.06%, therefore chloride was not a factor in the soda ash makeup. Daily coal analysis is included in Appendix A.

Table 4.5

Acceptance Test Daily Average Filter Cake Analysis

Test Day	As Received Basis			Insoluble Solids wt %	Mole Na_2CO_3 / Mole SO_2	Mole CaO / Mole SO_2
	Na wt %	Ca wt %	Total Sulfur as SO_4 wt %			
1	0.55	14.88	31.35	52.65	0.037	1.139
2	0.58	14.60	31.58	52.20	0.038	1.109
3	0.70	15.64	32.60	52.60	0.045	1.151
4	0.48	15.72	32.68	53.72	0.031	1.154
5	0.45	15.15	31.92	53.92	0.029	1.139
6	1.11	15.20	33.20	51.90	0.070	1.099
7	0.58	14.35	31.80	50.70	0.038	1.083
8	0.50	14.44	32.32	51.40	0.032	1.072
9	1.07	14.05	32.58	51.00	0.068	1.035
10	0.45	14.43	33.07	52.43	0.028	1.047
11	0.77	14.85	33.48	52.82	0.047	1.064
12	0.62	13.74	31.64	50.58	0.041	1.042
Average				52.16	0.042	1.095

Calcium consumption corrected for available alkalinity (1.095 x 0.95* = 1.040)

* Correction factor developed from analysis of incoming carbide lime for mole of available alkalinity per mole of total calcium

4.1.5 Power Consumption

"At the peak operating rate (300 MW), the [FGD] system will consume a maximum of 1.2% of the power generated by the unit."

This guarantee is interpreted as a maximum power consumption by the D/A FGD system of 3.6 MW, excluding the energy of the fuel oil burned in the reheater or power consumed during start-up. The power consumption guarantee may be exceeded by 30% during Acceptance Testing, but must be met during the Demonstration year.

The gross power generated by the unit and the energy consumed by the FGD system were recorded on separate watt-hour meters. During the Acceptance Test the peak generation was 240 megawatts (MW). Correspondingly, the power consumed during peak generation was 2.5 MW, or 1.05%. The guarantee was met based on peak generation and also based on average generation over the test period. During the 12-day test the average load was 178 MW and the average power consumption by the FGD system was 2.05 MW, or 1.15%. Appendix A also includes data on daily gross power generated and power consumed.

4.1.6 Waste Solids Properties

"The waste [filter cake] produced by the vacuum filter will contain a minimum of 55% insoluble solids".

The maximum level of moisture in the cake may exceed 30% during Acceptance Testing, but must be met during the Demonstration Year".

The filter cake averaged 52.2 wt % insoluble solids during the Acceptance Test. While this fell slightly short of guarantee, it was within the requirements of the Acceptance Test. Also, the product discharged to the IUCS process was uniform in moisture content and was suitable for working into a stable and manageable product through the fixative process. Optimization of filter cloth selection and filter cycle will continue with goal of showing that compliance with this guarantee can be met during the Demonstration year. Appendix A contains the results of moisture analysis of all samples taken during the test period.

4.1.7 System Availability

"The [FGD] system will have an availability (as defined by the Edison Electric Institute for power plant equipment) of at least 90% for one year [the Demonstration year]."

The Edison Electric Institute defines availability as follows:

"OPERATING AVAILABILITY - The percent of time the unit was available for service, whether operated or not. It is equal to available hours divided by the total hours in the period under consideration expressed as a percentage.

AVAILABLE - The status of a major piece of equipment which is capable of service, whether or not it is actually in service."

While it is too early to report the yearlong availability at this time, the availability through seven months of the Demonstration Year (May-November 1980) has averaged 96.9%. Table 4.6 summarizes the reliability

and availability of the D/A FGD system through November. Appendix B presents the daily summary of boiler and FGD performance parameters through the same period.

Table 4.6

Demonstration Year System Performance May 1980 Through November 1980

Period Hours (A)	5137.0
Boiler Hours (B)	4455.2
Boiler Utilization (B/A),%	86.73
FGD Hours (C)	4391.2
Hours FGD Available (D)	4977.8
Hours FGD Called Upon (E)	4451.0
FGD Availability (D/A),%	96.90
FGD Reliability (C/E),%	98.66
FGD Operability (C/B),%	98.56
FGD Utilization (C/A),%	85.48

4.2 Test Program

The Acceptance Test began at 0000 hour July 17, 1980 immediately after the boiler came back on line after a shutdown to repair a boiler tube leak. The program ran smoothly for the first six days except for the discrepancy between the Lear Siegler readings and the EPA Method 6 analysis. This was discovered after preliminary analysis from the first day of stack testing was completed. After a Lear Siegler representative arrived on site and a thorough check of the system was completed, a damaged ground wire was discovered. The effect on the analyzer was that the meter still tracked changes of SO_2 levels, but the recorded value was consistently lower than Method 6 analysis. This was verified when the pH in the absorber dropped below 6.0 as a result of a pH probe malfunction. The Lear Siegler analy-

zer followed the corresponding rise in SO_2 emissions. Careful analysis of calibration sequences before and after the burned ground was repaired showed the analyzer results had been low by 30 ppm. The Lear Siegler results for the first six days were therefore corrected by adding 30 ppm to the hourly and daily averages.

For each test day, the boiler load was raised to its maximum achievable level (240-250 MW) some time between 0600 and 0800 hours. The load was limited to this level because the FGD system booster fan could handle no more gas volume. The boiler was being operated with an absorber inlet oxygen concentration of 7.7% instead of the design of 5.9.%. This resulted in a larger volume of gas for the corresponding generation rate. The volumetric flowrate as determined by the pitot traverses by Kenvirons showed the system to be at maximum design gas flowrate (1,000,000 acfm), yet only at 80% of design generating capacity.

High load was held daily for 10 hours (an Acceptance Test prerequisite) and then reduced in steps to a low level of about 80 MW for about 6-7 hours generally between 0000-0700 hours in line with electrical energy demand.

The test proceeded smoothly until July 23 when wet coal resulting from heavy rains during the previous night blocked the coal feeders to the mills. The boiler load was restricted to 150-160 MW for two days. Process stream analysis and stack sampling continued because the FGD system was performing well, but it was decided to extend the test program by two days to record 10 days of operation at full capacity.

By July 25 the wet coal problems subsided, and the program continued to completion at 2400 hours on July 28.

4.3 Methodology

The data necessary to evaluate the performance of the D/A system was collected from two sources; the process instrumentation and the analysis of representative samples extracted from the operating streams. The flow recording and process analytical instrumentation furnished as part of the CEA/ADL design was calibrated prior to the start of the test.

Kenvirons performed daily combination EPA Method 5-8 tests [Reference (2)] to measure particulate matter emissions and SO_2 simultaneously at the stack and absorber inlets. The former as the only measure of the particulate matter guarantee and the latter as an ongoing check of and backup to the continuous SO_2 and O_2 analyzers. Corresponding oxygen and moisture measurements were performed using EPA Methods 3 and 4.

Samples from the process streams were collected during periods of steady state operation, where this was not possible, representative composite samples were obtained. The analytical techniques that were used during the Acceptance Test are summarized in Table 4.7. The specific details of each of these procedures are incorporated in the Dual Alkali Laboratory Manual prepared by Bechtel [Reference (3)].

4.3.1 Testing Procedures

Sample Locations - The stack emissions were monitored continuously by a Lear Siegler Model SM810 for SO_2 and a Lear Siegler CM50 for O_2.

Table 4.7

Analytical Methods

Liquor Analyses	Method	
Determination	Primary	Backup
Calcium	Ion Chromatograph	EDTA Titration
Magnesium	"	EDTA Titration
Sodium	"	Specific Ion Electrode
Sulfate	"	$BaSO_4$ Turbidimetric
Total Oxidizable Sulfur (TOS)	"	I_2/Thio Titration
Chloride	"	$Hg(NO_3)_2$ Titration
Fluoride	"	Specific Ion Electrode
Nitrate	"	Chromotropic Acid
Thiosulfate	"	I_2/Thio Titration
Alkalinity	HCl Titration	
Hydroxide	" "	
Sulfite	I_2/Thio Titration	
Dissolved Solids	Gravimetry or Calculation	
Trace Metals	A/A (by others)	

Solid Analyses		
% HCl Insol.	Gravimetry	
Suspended Solids	"	
Alkalinity	HCl/NaOH Titration	
Carbonate	Volumetric - CO_2 evolution	
Hydroxide	HCl Titration	
Sulfite (TOS)	I_2/Thio Titration	
Particle Size Distribution	Sieves/Sub-sieve analysis	
Calcium	Ion Chromatograph	EDTA Titration
Magnesium	"	EDTA Titration
Sodium	"	Specific Ion Electrode
Sulfate	"	$BaSO_4$ Turbidimetric
Nitrate	"	Chromotropic Acid
Chloride	"	$Hg(NO_3)_2$ Titration
Trace Metals	A/A (by others)	

Inlet SO_2 emissions were measured by single DuPont 460 ultraviolet analyzer which alternatively draws sample gas from each side every 10 minutes. A portable Thermox O_2 analyzer was used to manually check O_2 concentration from each side once every hour.

All testing was conducted at three locations in the ductwork between the ESP and stack. Figure 4.1, a plan view of the ESP, scrubber, and stack shows the location of the analyzers and the test samples ports. The gas stream is split into two equal parts as it leaves the boiler. Thus, the ESP and scrubber consist of an "A" side (north) and "B" side (south) as illustrated. The ducts are connected to a common plenum after the ESP (before scrubber) to allow pressure balancing. Also, this plenum serves as a scrubber bypass duct. The three test locations are:

1. ESP Outlet "A"
2. ESP Outlet "B"
3. Stack

EPA Method 1 was used as a general guide in selecting the number and location of sample points. At the ESP location, compromises were made due to the lack of straight ductwork prior to ports. Generally, a minimum of 2 minutes per point and approximately 50 points per duct for non-Method 2 locations was targeted. The sample representative area per point (length and width) dimension ratio did not exceed 2:1. The sample time per point was adjusted to allow a total time at all three locations for the day to be approximately simultaneous. Table 4.8 presents a summary of test location characteristics.

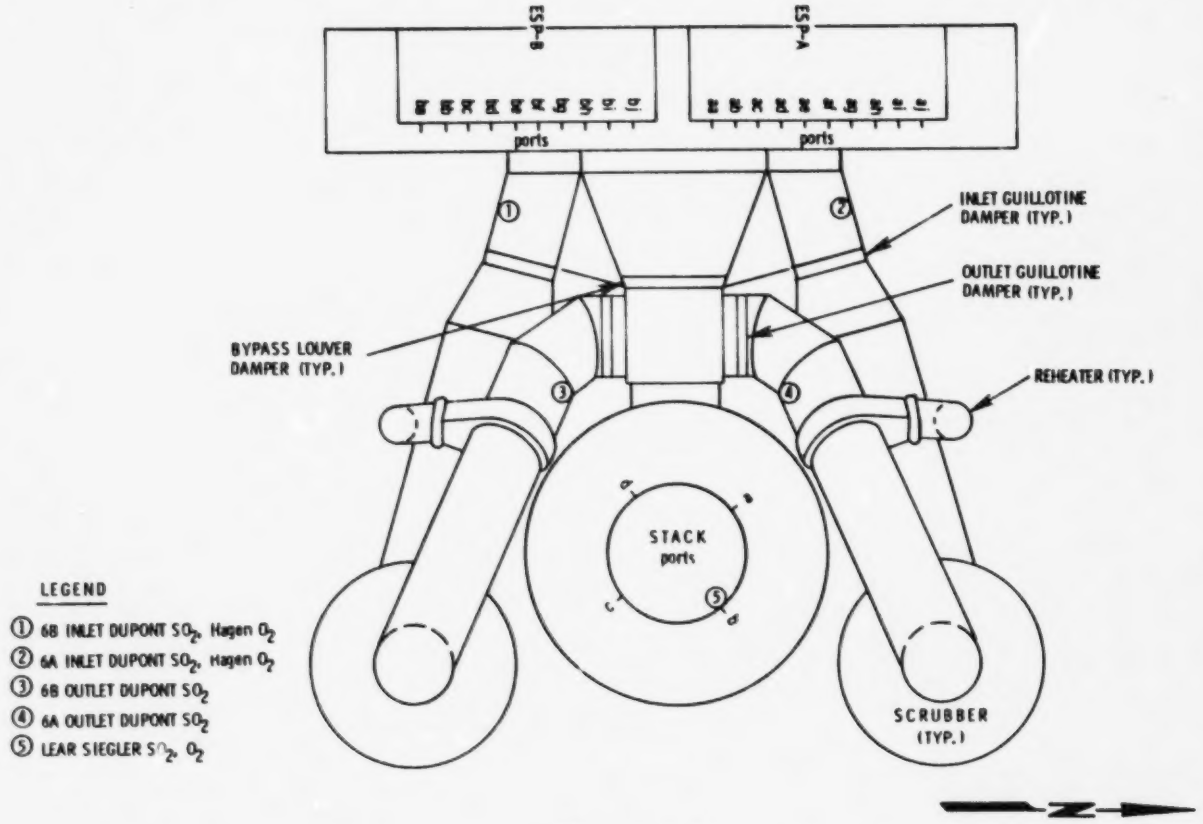

Figure 4.1

Sample Locations

Table 4.8

Sample Location	No. Ports	Points Per Port	Total Points	Minutes Per Point	Total Time (minutes)
ESP Outlet A	10	5	50	2	100
ESP Outlet B	10	5	50	2	100
Stack	4	7	28	3	84

4.3.2 Determination of Results

SO_2 Emissions - The primary method for determining SO_2 removal efficiency relied on the continuous SO_2 analyzers at the inlet to each absorber and in the stack. The analyzers are not equipped to remove water vapor from the gas stream prior to the analyzer. Therefore, necessary adjustments to the analyzer readings were made to correct dry conditions according to the results of Kenvirons testing. Volume II and III contain the details of Kenvirons test results and calculation procedures.

The average SO_2 readings were obtained from the electronic integrators installed on each of the analyzers. Daily SO_2 emissions in and out of the system were calculated in pounds per million Btu from the analyzer read-out using the F-factor technique outlined in EPA Method 19. The 200 ppm standard was determined by correcting the daily average stack SO_2 readings for the effect of the reheater and adjusted to "normal air dilution" as follows:

$$SO_2 \text{ (Corrected)} = SO_2 \text{ (Stack)} \times [20.9 - O_2 \text{ (Inlet)}]/[20.9-O_2 \text{ (Outlet)}]$$

Where:

O_2 (Inlet) = Absorber inlet oxygen concentration

O_2 (Outlet) = Stack oxygen concentration

Particulate Emissions - The particulate emission guarantee was checked by measuring the particulate loading in the stack simultaneously with measurements at the inlet of both the scrubbers. Volumes II and III of this report contain details of Kenvirons test results and calculation procedures. The particulate emission rate for the stack was calculated as follows:

$$PME = PML \text{ [1/7000 lb/grain]} \times F \times [20.9]/[20.9-\%O_2 \text{ (Outlet)}]$$

Where:

PME = Particulate Mass Emission Rate (lb/MM Btu)

PML = Particulate Mass Loading (grain/dscf)

F = Factor outlined in EPA Method 19 (dscf/10^6 Btu)

Lime Consumption - The lime stoichiometry was determined by analyzing representative samples of filter cake collected as the cake was discharged from the filters prior to fixation. The cake was analyzed for total calcium and total sulfur. The total calcium represented the lime used, and the total sulfur represented SO_2 removed from the flue gas. A portion of the calcium entering the system with the carbide lime is present as carbonate and therefore does not represent alkalinity available for regeneration. Each time the lime day tank was filled, a sample

of lime was analyzed for total calcium and for available alkalinity de- as an HCl titration to a pH of 10.2. From these results, a correction factor was developed to account for unreactive calcium in the carbide lime feed.

Over the test period a net lime stoichiometry was calculated by averaging the filter cake analyses over 12 days with a 12-day average feed analysis correction for available alkalinity as follows:

$$\text{Net Lime Stoichiometry} = (\text{Avg. } Ca^{++}/\text{Avg. S})_{\text{Cake}} \times (\text{Avg. Alk}/\text{Avg. } Ca^{++})_{\text{Lime}}$$

An example of calculation of lime and soda ash consumption is given on page A-13.

Soda Ash Consumption - The soda ash consumption was determined from the feed streams and from the filter cake in the same manner as the lime consumption except there was no need for correction of available soda ash.

$$\text{Net Soda Ash Stoichiometry} = (\text{Avg. } Na+/(\text{Avg. S})_{\text{Cake}}$$

Power Consumption - Over the 12-day period, the energy consumption by the FGD system was compared to the gross electrical generation of Unit 6. Using kilowatt-hour meters to measure total electrical energy supplied to the dual alkali system and gross turbine production by Unit 6.

Filter Cake Solids - The moisture content and percent insolubles was determined for each filter cake sample. A daily average of percent insoluble solids was determined for monitoring purposes. An average over the 12-day period was used for determining guarantee performance.

4.4 Quality Assurance

All samples for analysis during the Acceptance Test were logged in, analyzed and reported according to the procedures outlined in the published laboratory manual [Reference (3)]. Representative sample analyses were duplicated and also spiked. The results were compared to Shewhart charts established prior to the Acceptance Test. A round robin laboratory analyses program was created to evaluate the quality and accuracy of the analyses performed at the Dual Alkali Laboratory. Laboratories participated in the round robin tests were: Shawnee, Accurex, Bechtel, TVA Muscle Shoals, Kenvirons, and Dual Alkali. Overall, the Dual Alkali Laboratory performed well in these tests.

Kenvirons followed a quality assurance program which involved calibration of sampling instrumentation, standardization of reagents, duplicate and spiked analyses, and participation in round robin laboratory analyses. Details of their quality assurance procedures are given in Volume I and II of this report.

REFERENCES

1. Van Ness, R.P., et. al., "Full-Scale Dual Alkali Demonstration System at Louisville Gas and Electric Company - Final Design and System Cost", EPA 600/7-79-221b, NTIS No. PB80 146715, September 1979.

2. 40 CFR 60.1, Code of Federal Regulation, August 18, 1977.

3. Donnelly, J.R., et. al., "Laboratory Procedures For Analysis of Sodium Based Dual Alkali Process Streams", EPA-600/8-80-015, March 1980.

Appendix A

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 1 Date: July 17, 1980

Filter cake removed, (tons/day)	-	337.3
Soda ash added, (tons/day)	-	16.3
Carbide lime added, (tons/day)	-	735.9
Gross energy generated, (MW-hr)	-	4,241.0
Peak power generated, (MW)	-	238.0
Gross energy consumed by FGD, (MW-hr)	-	48.0
Power consumed by FGD at peak load, (MW)	-	2.5
Coal burned, (tons/day)	-	31,884.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2234	2210	81.5

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	61.3	14.3	4.5	6.6	1.2	3.4	.01	8.7
	59.7	14.0	4.3	7.4	1.2	3.9	.01	9.4

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	2340	14.6	0.7	30.3	50.3	46.9
	A	2350	15.1	0.4	32.4	55.0	43.2

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0130	11.6	19.9	22.8
	1445	11.8	21.3	23.8

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 2 Date: July 18, 1980

Filter cake removed, (tons/day)	-	762.8
Soda ash added, (tons/day)	-	18.4
Carbide lime added, (tons/day)	-	527.4
Gross energy generated, (MW-hr)	-	4,285.0
Peak power generated, (MW)	-	241.0
Gross energy consumed by FGD, (MW-hr)	-	50.4
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	32,049.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2444	2349	80.3

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	61.6	14.5	4.3	6.6	1.3	3.6	.01	8.1

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0330	15.2	0.6	33.7	50.9	42.9
	A	0340	14.1	0.5	30.1	51.6	48.6
	B	0815	14.4	0.5	31.1	50.9	47.6
	A	0825	13.7	0.7	31.9	52.6	46.0
	B	1640	15.7	0.6	32.8	55.0	44.1

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0000	12.1	21.7	24.7
	0940	11.9	21.1	23.9

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 3 Date: July 19, 1980

Filter cake removed, (tons/day)	-	621.4
Soda ash added, (tons/day)	-	15.4
Carbide lime added, (tons/day)	-	685.9
Gross energy generated, (MW-hr)	-	4,179.0
Peak power generated, (MW)	-	242.0
Gross energy consumed by FGD, (MW-hr)	-	47.4
Power consumed by FGD at peak load, (MW)	-	2.5
Coal burned, (tons/day)	-	31,838.0

Sulfur dioxide	Abs. SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2231	2185	81.5

Coal analysis As burned, (wt %)

C	Ash	H	O	N	S	Cl	H_2O
63.4	14.9	4.4	6.9	1.2	3.9	.01	5.3

Filter cake analysis: As received, wt %

Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
B	0800	16.1	0.4	32.8	53.5	45.6
A	2050	15.2	1.0	32.4	51.7	45.8

Lime slurry analysis: As received, (wt %)

Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
0455	12.6	22.1	24.8
1700	12.5	22.1	24.6

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 4 Date: July 20, 1980

Filter cake removed, (tons/day)	-	545.0
Soda ash added, (tons/day)	-	17.0
Carbide lime added, (tons/day)	-	459.0
Gross energy generated, (MW-hr)	-	4,266.0
Peak power generated, (MW)	-	242.0
Gross energy consumed by FGD, (MW-hr)	-	49.0
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	30,527.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2222	2093	99.7

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	61.0	15.2	4.3	6.5	1.3	4.2	.01	7.4

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	A	0110	16.1	0.6	33.6	53.9	45.4
	B	0405	16.2	0.4	33.7	53.5	45.7
	B	1100	16.0	0.4	32.4	53.4	46.0
	A	1113	15.6	0.5	32.9	54.6	45.4
	B	1600	14.6	0.4	31.3	53.0	46.4
	A	1610	15.8	0.6	32.2	53.9	44.9

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0400	12.6	21.7	24.0
	1620	12.1	21.6	24.3

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 5 Date: July 21, 1980

Filter cake removed, (tons/day)	-	580.0
Soda ash added, (tons/day)	-	14.9
Carbide lime added, (tons/day)	-	NA
Gross energy generated, (MW-hr)	-	4,323.0
Peak power generated, (MW)	-	241.0
Gross energy consumed by FGD, (MW-hr)	-	49.6
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	32,590.0

Sulfur dioxide	Abs. SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2191	2106	103.8

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	63.0	12.6	4.4	6.2	1.3	3.7	.01	8.8

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0345	14.2	0.45	31.8	52.4	47.2
	B	1320	14.3	0.5	30.8	53.5	45.8
	B	1700	15.6	0.45	31.0	53.5	45.4
	A	1705	16.5	0.4	34.0	56.3	42.7

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0340	12.0	21.8	24.6
	1650	12.1	21.7	24.3

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 6 Date: July 22, 1980

Filter cake removed, (tons/day)	-	240.7
Soda ash added, (tons/day)	-	9.0
Carbide lime added, (tons/day)	-	509.8
Gross energy generated, (MW-hr)	-	4,444.0
Peak power generated, (MW)	-	242.0
Gross energy consumed by FGD, (MW-hr)	-	47.9
Power consumed by FGD at peak load, (MW)	-	2.4
Coal burned, (tons/day)	-	32,809.0

Sulfur dioxide	Abs. SO_2 in	Abs. SO_2 in	Stack SO_2
Daily average, (ppm)	2346	2299	89.7

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	60.1	12.2	4.3	6.6	1.2	3.7	.04	12.2

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0130	15.2	1.9	33.2	51.9	43.3

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0445	12.0	21.2	23.8
	1700	11.0	19.5	22.3

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 7 Date: July 23, 1980

Filter cake removed, (tons/day)	-	636.1
Soda ash added, (tons/day)	-	4.8
Carbide lime added, (tons/day)	-	655.4
Gross energy generated, (MW-hr)	-	4,266.0
Peak power generated, (MW)	-	231.0
Gross energy consumed by FGD, (MW-hr)	-	48.4
Power consumed by FGD at peak load, (MW)	-	2.4
Coal burned, (tons/day)	-	30,682.0

Sulfur dioxide	Abs. SO_2 in	Abs. SO_2 in	Stack SO_2
Daily average, (ppm)	1931	1847	83.3

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	58.8	13.2	4.2	6.5	1.4	3.3	.04	12.5

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0015	13.7	0.4	30.8	49.3	49.8
	A	0020	13.6	0.4	30.8	48.3	50.3
	C	1250	15.8	0.4	32.8	53.9	45.4
	B	1300	14.9	0.8	32.0	51.3	47.2
	C	1800	13.7	1.6	32.0	NA	47.4
	B	1810	14.4	0.9	32.8	NA	46.9

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	1625	11.6	18.9	21.0
	0400	11.8	20.1	22.5

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 8 Date: July 24, 1980

Filter cake removed, (tons/day)	-	561.6
Soda ash added, (tons/day)	-	3.9
Carbide lime added, (tons/day)	-	606.2
Gross energy generated, (MW-hr)	-	4,437.0
Peak power generated, (MW)	-	242.0
Gross energy consumed by FGD, (MW-hr)	-	51.5
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	33,847.0

Sulfur dioxide	Abs. SO_2 in	Abs. SO_2 in	Stack SO_2
Daily average, (ppm)	1934	1909	106.4

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	62.0	12.1	4.4	7.0	1.3	3.2	.05	9.0
	59.8	13.0	4.3	6.4	1.4	3.2	.00	12.0

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0000	14.8	0.4	31.8	NA	47.4
	C	0010	14.5	0.45	31.1	NA	48.5
	B	1015	14.1	0.75	32.9	51.4	46.2
	C	1640	14.7	0.4	33.6	55.1	43.5
	B	1650	14.1	1.5	31.9	51.1	47.3

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0350	11.2	18.5	21.2
	1655	11.6	18.9	22.9

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 9 Date: July 25, 1980

Filter cake removed, (tons/day)	-	502.7
Soda ash added, (tons/day)	-	3.6
Carbide lime added, (tons/day)	-	783.9
Gross energy generated, (MW-hr)	-	4,392.0
Peak power generated, (MW)	-	243.0
Gross energy consumed by FGD, (MW-hr)	-	50.6
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	35,031.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack O_2
Daily average, (ppm)	2189	2138	124.6

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	65.5	14.3	4.6	7.1	1.6	3.7	.05	3.1

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0510	14.2	1.2	32.2	52.3	46.1
	A	0515	14.7	0.7	32.9	52.7	45.9
	B	0820	13.9	0.8	32.3	50.8	47.5
	A	0825	14.4	1.5	34.5	51.6	44.3
	A	1700	14.0	0.95	32.5	51.8	46.1
	C	1705	13.1	1.3	31.1	46.8	47.8

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0500	10.7	18.8	21.9
	1650	11.7	20.8	23.4

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 10 Date: July 26, 1980

Filter cake removed, (tons/day)	-	515.1
Soda ash added, (tons/day)	-	3.3
Carbide lime added, (tons/day)	-	714.1
Gross energy generated, (MW-hr)	-	4,202.0
Peak power generated, (MW)	-	240.0
Gross energy consumed by FGD, (MW-hr)	-	49.2
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	31,390.0

Sulfur dioxide	Abs. A SO_2 in	Abs. SO_2 in	Stack SO_2
Daily average, (ppm)	2168	2116	145.9

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	60.2	14.7	4.2	5.4	1.4	4.7	.04	9.5

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	A	0015	14.6	0.5	33.2	53.5	46.3
	B	1605	14.1	0.3	32.4	51.3	48.0
	A	1610	14.6	0.55	33.6	52.5	46.4

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0435	12.0	20.6	23.6
	1615	11.6	20.7	23.0

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 11 Date: July 27, 1980

Filter cake removed, (tons/day)	-	487.0
Soda ash added, (tons/day)	-	4.9
Carbide lime added, (tons/day)	-	661.0
Gross energy generated, (MW-hr)	-	4,157.0
Peak power generated, (MW)	-	240.0
Gross energy consumed by FGD, (MW-hr)	-	49.5
Power consumed by FGD at peak load, (MW)	-	2.6
Coal burned, (tons/day)	-	32,027.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	2095	2041	133.7

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	58.2	14.0	4.2	6.3	1.3	3.6	.00	12.4

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0020	14.8	0.3	32.0	51.7	46.4
	A	0030	14.7	0.3	33.1	52.9	46.6
	B	2140	14.6	1.5	34.2	51.7	43.2
	A	2150	15.3	1.0	34.6	55.0	42.3

Lime slurry analysis: As received, (wt %)	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0400	11.6	20.4	23.0

*Loss on Drying

ACCEPTANCE TEST DAILY SUMMARY

Acceptance Test Day: 12 Date: July 28, 1980

Filter cake removed, (tons/day)	-	608.9
Soda ash added, (tons/day)	-	8.1
Carbide lime added, (tons/day)	-	477.6
Gross energy generated, (MW-hr)	-	4,021.0
Peak power generated, (MW)	-	236.0
Gross energy consumed by FGD, (MW-hr)	-	47.7
Power consumed by FGD at peak load, (MW)	-	2.5
Coal burned, (tons/day)	-	31,915.0

Sulfur dioxide	Abs. A SO_2 in	Abs. B SO_2 in	Stack SO_2
Daily average, (ppm)	1981	1980	111.3

Coal analysis As burned, (wt %)	C	Ash	H	O	N	S	Cl	H_2O
	61.2	13.3	4.3	6.5	1.4	3.5	.01	9.8
	58.4	14.3	4.2	7.0	1.3	3.4	.00	11.4

Filter cake analysis: As received, wt %	Filter	Time	Ca	Na	Total S as SO_4	Insol. Solids	LOD*
	B	0500	14.1	0.5	31.9	50.6	46.7
	A	0510	14.1	0.7	32.7	51.5	46.1
	B	1325	13.5	0.5	30.5	50.3	49.2
	C	1700	12.7	0.9	30.4	47.0	50.0
	B	1705	14.3	0.3	32.7	53.5	44.6

Lime slurry analysis:	Time	Ca	Avail. Alk as $Ca(OH)_2$	Insol. Solids
	0400	12.6	NA	25.0

*Loss on Drying

SAMPLE CALCULATION FOR LIME AND SODA ASH CONSUMPTION

Filter Cake Analysis (wt %)

Ca: 14.6, Na:0.7, Total S (as SO_4): 30.3

Lime Slurry Analysis (as received - wt %)

Ca: 11.6, Available Alkalinity (as $Ca(OH)_2$): 19.9

Lime Consumption: $(\text{Avg. } Ca^{++}/\text{Avg. S})_{\text{Cake}} \times (\text{Avg. Alk/Avg. } Ca^{++})_{\text{Lime}}$

$$\frac{(14.6/40)}{(30.3/96)} \times \frac{(19.9/74)}{(11.6/40)} = 1.07 \text{ moles Lime/mole } SO_2 \text{ removed}$$

Soda Ash Consumption:

$$\frac{(0.7/(23x2))}{(30.3/96)} = 0.048 \text{ moles Soda Ash/mole } SO_2 \text{ removed}$$

Molecular Weights:

Ca - 40	SO_4 - 96
Na - 23	$Ca(OH)_2$ - 74

Appendix B. Monthly Performance Parameters

OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
05/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/02	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/03	23.5	23.5	24.0	23.5	097.92	097.92	100.00	100.00	100.00
05/04	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/06	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/07	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/08	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/09	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/10	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/11	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58
05/12	24.0	23.5	23.6	23.8	100.00	097.92	098.33	098.74	097.92
05/13	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
05/14	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/15	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/16	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/18	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/19	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58
05/20	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/21	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/22	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/23	21.5	21.5	24.0	21.5	089.58	089.58	100.00	100.00	100.00
05/24	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/25	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/26	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/27	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/28	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
05/29	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/30	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
05/31	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	744.0
BOILER HOURS (B)-	741.0
BOILER UTILIZATION (B/A),%-	099.60
FGD HOURS (C)-	740.0
HOURS FGD AVAILABLE (D)-	743.4
HOURS FGD CALLED UPON (E)-	740.5
FGD AVAILABILITY (D/A),%-	099.92
FGD RELIABILITY (C/E),%-	099.93
FGD OPERABILITY (C/B),%-	099.87
FGD UTILIZATION (C/A),%-	099.46

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OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGI UTILIZ. %	FGI AVAIL. %	FGI RELIAB. %	FGI OPER. %
06/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/02	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/03	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/04	24.0	23.7	23.7	24.0	100.00	098.75	098.75	098.75	098.75
06/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/06	23.1	23.1	24.0	23.1	096.25	096.25	100.00	100.00	100.00
06/07	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
06/08	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
06/09	23.6	23.6	24.0	23.6	098.33	098.33	100.00	100.00	100.00
06/10	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/11	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/12	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/13	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/14	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/15	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/16	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/18	24.0	23.0	24.0	23.0	100.00	095.83	100.00	100.00	095.83
06/19	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17
06/20	24.0	23.6	23.8	23.9	100.00	098.33	099.17	098.74	098.33
06/21	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/22	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/23	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/24	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/25	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/26	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/27	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17
06/28	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
06/29	19.2	19.2	24.0	19.2	080.00	080.00	100.00	100.00	100.00
06/30	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	720.0
BOILER HOURS (B)-	641.9
BOILER UTILIZATION (B/A),%-	089.15
FGD HOURS (C)-	639.8
HOURS FGI AVAILABLE (D)-	719.5
HOURS FGD CALLED UPON (E)-	640.4
FGD AVAILABILITY (D/A),%-	099.93
FGD RELIABILITY (C/E),%-	099.91
FGD OPERABILITY (C/B),%-	099.67
FGD UTILIZATION (C/A),%-	088.86

OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
07/01	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
07/02	14.3	13.8	24.0	13.8	059.58	057.50	100.00	100.00	096.50
07/03	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/04	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/06	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/07	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/08	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58
07/09	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17
07/10	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58
07/11	01.2	01.2	24.0	01.2	005.00	005.00	100.00	100.00	100.00
07/12	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
07/13	18.6	18.5	23.9	18.6	077.50	077.08	099.58	099.46	099.46
07/14	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
07/15	00.3	00.3	24.0	00.3	001.25	001.25	100.00	100.00	100.00
07/16	07.9	07.9	24.0	07.9	032.92	032.92	100.00	100.00	100.00
07/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/18	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/19	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/20	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/21	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/22	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/23	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/24	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
07/25	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/26	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/27	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/28	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/29	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/30	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
07/31	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	744.0
BOILER HOURS (B)-	618.3
BOILER UTILIZATION (B/A).%-	083.10
FGD HOURS (C)-	616.9
HOURS FGD AVAILABLE (D)-	743.5
HOURS FGD CALLED UPON (E)-	617.4
FGD AVAILABILITY (D/A).%-	099.93
FGD RELIABILITY (C/E).%-	099.92
FGD OPERABILITY (C/B).%-	099.77
FGD UTILIZATION (C/A).%-	082.92

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OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
08/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/02	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/03	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/04	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/06	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
08/07	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
08/08	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/09	00.4	00.4	24.0	00.4	001.67	001.67	100.00	100.00	100.00
08/10	19.0	00.0	24.0	19.0	079.17	000.00	100.00	000.00	000.00
08/11	24.0	12.7	24.0	24.0	100.00	052.92	100.00	052.92	052.92
08/12	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/13	19.9	19.9	24.0	19.9	082.92	082.92	100.00	100.00	100.00
08/14	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
08/15	01.2	01.2	24.0	01.2	005.00	005.00	100.00	100.00	100.00
08/16	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/18	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/19	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/20	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/21	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/22	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/23	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
08/24	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/25	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/26	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/27	24.0	19.4	19.4	24.0	100.00	080.83	080.83	080.83	080.83
08/28	24.0	23.7	23.7	24.0	100.00	098.75	098.75	098.75	098.75
08/29	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
08/30	24.0	23.9	24.0	23.8	100.00	099.58	100.00	100.42	099.58
08/31	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	744.0
BOILER HOURS (B)-	664.5
BOILER UTILIZATION (B/A),%-	089.3
FGD HOURS (C)-	628.5
HOURS FGD AVAILABLE (D)-	738.6
HOURS FGD CALLED UPON (E)-	664.1
FGD AVAILABILITY (D/A),%-	099.2
FGD RELIABILITY (C/E),%-	094.64
FGD OPERABILITY (C/B),%-	094.58
FGD UTILIZATION (C/A),%-	084.48

OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
09/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/02	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17
09/03	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
09/04	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/06	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/07	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/08	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/09	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/10	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/11	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/12	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/13	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/14	24.0	16.3	24.0	24.0	100.00	067.92	100.00	067.92	067.92
09/15	24.0	23.9	24.0	23.9	100.00	099.58	100.00	100.00	099.58
09/16	24.0	23.7	23.9	23.8	100.00	098.75	099.58	099.58	098.75
09/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/18	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/19	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/20	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/21	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/22	23.2	23.2	24.0	23.2	096.67	096.67	100.00	100.00	100.00
09/23	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
09/24	00.0	00.0	00.0	00.0	000.00	000.00	000.00	XXXXXX	XXXXXX
09/25	00.0	00.0	00.3	00.0	000.00	000.00	022.09	XXXXXX	XXXXXX
09/26	12.1	12.1	24.0	12.1	050.42	050.42	100.00	100.00	100.00
09/27	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/28	24.0	23.7	23.9	23.8	100.00	098.75	099.58	099.58	098.75
09/29	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
09/30	24.0	23.8	24.0	23.8	100.00	099.17	100.00	100.00	099.17

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	720.0
BOILER HOURS (B)-	693.3
BOILER UTILIZATION (B/A).%-	096.24
FGD HOURS (C)-	682.3
HOURS FGD AVAILABLE (D)-	676.9
HOURS FGD CALLED UPON (E)-	694.4
FGD AVAILABILITY (D/A).%-	094.01
FGD RELIABILITY (C/E).%-	098.72
FGD OPERABILITY (C/B).%-	098.58
FGD UTILIZATION (C/A).%-	094.76

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OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
11/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/02	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
11/03	24.0	18.9	18.9	24.0	100.00	078.75	078.75	078.75	078.75
11/04	24.0	23.9	24.0	24.0	100.00	099.58	100.00	099.58	099.58
11/05	24.0	20.6	20.6	24.0	100.00	085.83	085.83	085.83	085.83
11/06	24.0	23.9	24.0	24.0	100.00	099.58	100.00	099.58	099.58
11707	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/08	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/09	24.0	23.9	23.9	24.0	100.00	099.58	099.58	099.58	099.58
11/10	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/11	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/12	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/13	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/14	24.0	19.8	19.8	24.0	100.00	082.50	082.50	082.50	082.50
11/15	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/16	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
11/18	01.6	01.6	01.6	01.6	006.67	006.67	006.67	100.00	100.00
11/19	00.0	00.0	00.0	00.0	000.00	000.00	000.00	XXXXXX	XXXXXX
11/20	00.0	00.0	00.0	00.0	000.00	000.00	000.00	XXXXXX	XXXXXX
11/21	00.0	00.0	00.0	00.0	000.00	000.00	000.00	XXXXXX	XXXXXX
11/22	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/23	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/24	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/25	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/26	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/27	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/28	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/29	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX
11/30	00.0	00.0	24.0	00.0	000.00	000.00	100.00	XXXXXX	XXXXXX

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	720.0
BOILER HOURS (B)-	409.6
BOILER UTILIZATION (B/A),%-	056.89
FGD HOURS (C)-	396.5
HOURS FGD AVAILABLE (D)-	612.7
HOURS FGD CALLED UPON (E)-	409.6
FGD AVAILABILITY (D/A),%-	085.10
FGD RELIABILITY (C/E),%-	096.80
FGD OPERABILITY (C/B),%-	096.80
FGD UTILIZATION (C/A),%-	055.07

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OPERATING SUMMARY - 1980

DATE	BOILER HOURS	FGD HOURS	AVAIL. HOURS	CALL HOURS	BOILER UTILIZ. %	FGD UTILIZ. %	FGD AVAIL. %	FGD RELIAB. %	FGD OPER. %
10/01	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/02	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/03	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/04	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/05	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/06	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/07	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/08	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/09	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/10	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/11	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/12	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/13	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/14	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/15	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/16	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/17	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/18	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/19	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/20	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/21	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/22	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/23	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/24	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/25	24.0	23.8	23.8	24.0	100.00	099.17	099.17	099.17	099.17
10/26	25.0	25.0	25.0	25.0	100.00	099.17	099.17	099.17	099.17
10/27	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/28	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/29	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00
10/30	23.6	23.6	23.6	23.6	098.33	098.33	098.33	100.00	100.00
10/31	24.0	24.0	24.0	24.0	100.00	100.00	100.00	100.00	100.00

MONTHLY SYSTEM PERFORMANCE

PERIOD HOURS (A)-	745.0
BOILER HOURS (B)-	744.6
BOILER UTILIZATION (B/A),%-	099.95
FGD HOURS (C)-	743.2
HOURS FGD AVAILABLE (D)-	743.2
HOURS FGD CALLED UPON (E)-	744.6
FGD AVAILABILITY (D/A),%-	099.76
FGD RELIABILITY (C/E),%-	099.81
FGD OPERABILITY (C/B),%-	099.81
FGD UTILIZATION (C/A),%-	099.76

Appendix C. Kenviron's Acceptance Test Analysis

OUTPUT DATE - 08/27/80
TIME - 14:30:49

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

EQUATIONS USED FOR PARTICULATE CALCULATIONS

1. VOLUME OF DRY GAS SAMPLED AT STANDARD CONDITIONS, DSCF

$$VMSTD = \frac{17.636 * VM * Y * (PBC + PM/13.6)}{TM + 459.69}$$

2. VOLUME OF WATER VAPOR AT STANDARD CONDITIONS, SCF

$$VWV = 0.04707 * VW$$

3. PERCENT MOISTURE IN STACK GAS

$$PMOS = \frac{100. * VWV}{VMSTD + VWV}$$

4. MOLE FRACTION OF DRY STACK GAS

$$MD = 1 - \frac{VWV}{VMSTD + VWV}$$

5. AVERAGE MOLECULAR WEIGHT OF DRY STACK GAS

$$MWD = (PCO2 * 0.440) + (PO2 * 0.320) + ((PN2 + PCO) * 0.280)$$

6. MOLECULAR WEIGHT OF STACK GAS

$$MW = MWD * MD + 18.015 * (1-MD)$$

7. STACK GAS VELOCITY AT STACK CONDITIONS, FPM

$$VS = 5129.4 * CP * DPS * SQRT((TS + 459.69)) * SQRT(1 / (PS * MW))$$

8. STACK GAS VOLUMETRIC FLOW AT STANDARD CONDITIONS, DRY BASIS, DSCFM

$$QS = \frac{0.12247 * VS * AS * MD * PS}{TS + 459.69}$$

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BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

9. STACK GAS VOLUMETRIC FLOW AT STACK CONDITIONS, ACFM

$$QA = \frac{QS * (TS + 459.69)}{17.636 * PS * MD}$$

10. PERCENT ISOKINETIC

$$PERI = \frac{1039.6 * (TS + 459.69) * VWSTD}{VS * TT * PS * MD * (DN ** 2)}$$

11. TOTAL PARTICULATE LOADING AT STANDARD CONDITIONS, GR/DSCF

$$CAN = 0.015432 * (MF / VMSTD)$$

11.A. TOTAL PARTICULATE LOADING AT STANDARD CONDITIONS, LB/DSCF

$$C = CAN / 7000.0$$

12. TOTAL PARTICULATE LOADING AT STACK CONDITIONS, GR/ACF

$$CAT = \frac{17.636 * CAN * PS * MD}{TS + 459.69}$$

13. TOTAL PARTICULATE EMISSION RATE, LBS/HR

$$CAW = 0.008571 * CAN * QS$$

14. EMISSION RATE, LB/MMBTU

$$E = C * F * 20.9 / (20.9 - PO2)$$

15. SO2 CONCENTRATION AT STANDARD CONDITIONS, LB/DSCF

$$SCONC = \frac{7.061E\text{-}5 * NBACL2 * (VT-VTB) * (VSOLN/VA)}{VMSTD}$$

16. SO2 CONCENTRATION AT STACK CONDITIONS, LB/ACF

$$SCONCA = \frac{17.636 * SCONC * PS * MD}{TS + 459.69}$$

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

17. SO2 EMISSION RATE, LB/HR

SCHR = 60.0 * SCONC * QS

18. SO2 EMISSION RATE, LB/MMBTU

SCMM = SCONC * F * (20.9 / (20.9 - PO2))

19. SO2 CONCENTRATION AT STANDARD (DRY) CONDITIONS, PPM (VOL.)

SCPPM = SCONC / (64.0 * 2.59E-9)

20. SO2 CONCENTRATION AT STACK (WET) CONDITIONS, PPM (VOL.)

SCPPMA= SCPPM * MD

OUTPUT DATE - 08/27/80
TIME - 14:30:51

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA01	ESPB01	STK001
	DATE OF RUN		07/17/80	07/17/80	07/17/80
	TIME OF RUN		1428-1724	1428-1655	1422-1702
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0240	1.0370	1.0170
PB	BAROMETRIC PRES, LAB	IN.HG	29.66	29.66	29.66
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.61	29.61	29.44
PM	AVG ORF PRES DROP	IN.H2O	0.580	0.706	0.532
VM	VOL DRY GAS-METER COND	DCF	42.60	48.70	36.18
TM	AVG GAS METER TEMP	DEG.F	104.8	126.4	114.0
VMSTD	VOL DRY GAS-STD COND	DSCF	40.41	45.07	33.34
VW	TOTAL H2O COLLECTED	ML	81.1	89.5	68.1
VWV	VOL H2O VAPOR-STD COND	SCF	3.82	4.21	3.21
PMOS	PERCENT MOISTURE BY VOLUME		8.6	8.5	8.8
MD	MOLE FRACTION DRY GAS		0.914	0.915	0.912
PCO2	PERCENT CO2 BY VOLUME, DRY		12.0	12.0	10.4
PO2	PERCENT O2 BY VOLUME, DRY		7.7	7.3	8.8
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.2
PN2	PERCENT N2 BY VOLUME, DRY		80.3	80.7	80.6
MWD	MOLECULAR WT-DRY STK GAS		30.23	30.21	30.02
MW	MOLECULAR WT-STK GAS		29.17	29.17	28.96
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.522	0.560	0.844
TS	AVG STACK TEMPERATURE	DEG.F	300.1	283.4	169.2
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.46	28.50	29.38
VS	AVG STACK GAS VELOCITY	FPM	2150	2280	3120
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	303000	329000	666000
QA	ACTUAL STACK FLOWRATE	ACFM	502000	532000	886000
PERI	PERCENT ISOKINETIC		100.7	100.9	104.7
MT	PARTICULATE WT-TOTAL	MG	628.70	962.90	79.80
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.240	0.330	0.037
C	PART. LOAD-TTL, STD	LB/DSCF	.00003430	.00004710	.00000527
CAT	PART. LOAD-TTL, STK	GR/ACF	0.145	0.204	0.028
CAW	PART. EMIS-TOTAL	LB/HR	624.0000	930.0000	211.0000
E	EMISSION RATE	LB/MMBTU	0.5320	0.7120	0.0895
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.86E-04	3.86E-04	2.05E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.33E-04	2.39E-04	1.54E-05
SCHR	SO2 EMISSION RATE	LB/HR	7028.138	7627.597	819.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.990	5.840	0.348
SCPPM	SO2, CONC., STD	PPM	2330	2330	124
SCPPMA	SO2, CONC., STK	PPM	2130	2130	113

OUTPUT DATE - 08/27/80
TIME - 14:31:50

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA02	ESPB02	STK002
	DATE OF RUN		07/18/80	07/18/80	07/18/80
	TIME OF RUN		1100-1338	1115-1342	1100-1301
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0250	1.0360	1.0175
PB	BAROMETRIC PRES, LAB	IN.HG	29.80	29.80	29.80
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.75	29.75	29.58
PM	AVG ORF PRES DROP	IN.H2O	0.640	0.647	0.577
VM	VOL DRY GAS-METER COND	DCF	44.11	45.99	37.82
TM	AVG GAS METER TEMP	DEG.F	104.1	125.1	112.4
VMSTD	VOL DRY GAS-STD COND	DSCF	42.14	42.81	35.14
VW	TOTAL H2O COLLECTED	ML	91.6	93.9	131.5
VWV	VOL H2O VAPOR-STD COND	SCF	4.31	4.42	6.19
PMOS	PERCENT MOISTURE BY VOLUME		9.3	9.4	15.0
MD	MOLE FRACTION DRY GAS		0.907	0.906	0.850
PCO2	PERCENT CO2 BY VOLUME, DRY		12.0	12.3	10.8
PO2	PERCENT O2 BY VOLUME, DRY		7.5	6.8	8.3
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		80.5	80.8	80.8
MWD	MOLECULAR WT-DRY STK GAS		30.22	30.25	30.06
MW	MOLECULAR WT-STK GAS		29.09	29.10	28.26
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.548	0.509	0.887
TS	AVG STACK TEMPERATURE	DEG.F	311.5	304.1	175.0
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.64	28.64	29.52
VS	AVG STACK GAS VELOCITY	FPM	2270	2100	3330
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	315000	294000	659000
QA	ACTUAL STACK FLOWRATE	ACFM	530000	490000	945000
PERI	PERCENT ISOKINETIC		101.0	107.4	111.6
MT	PARTICULATE WT-TOTAL	MG	822.10	482.50	90.90
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.301	0.174	0.040
C	PART. LOAD-TTL, STD	LB/DSCF	.00004300	.00002480	.00000570
CAT	PART. LOAD-TTL, STK	GR/ACF	0.179	0.104	0.028
CAW	PART. EMIS-TOTAL	LB/HR	813.0000	438.0000	225.0000
E	EMISSION RATE	LB/MMBTU	0.6590	0.3620	0.0932
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.57E-04	3.61E-04	2.71E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.12E-04	2.16E-04	1.89E-05
SCHR	SO2 EMISSION RATE	LB/HR	6752.409	6359.639	1069.728
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.470	5.260	0.442
SCPPM	SO2, CONC., STD	PPM	2150	2180	163
SCPPMA	SO2, CONC., STK	PPM	1950	1970	139

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OUTPUT DATE - 08/27/80
TIME - 14:32:23

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA03	ESPB03	STK003
	DATE OF RUN		07/19/80	07/19/80	07/19/80
	TIME OF RUN		1036-1302	1035-1307	1040-1231
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0250	1.0390	1.0158
PB	BAROMETRIC PRES, LAB	IN.HG	29.86	29.86	29.86
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.81	29.81	29.64
PM	AVG ORF PRES DROP	IN.H2O	0.652	0.790	0.487
VM	VOL DRY GAS-METER COND	DCF	45.59	51.15	34.77
TM	AVG GAS METER TEMP	DEG.F	107.0	116.9	112.2
VMSTD	VOL DRY GAS-STD COND	DSCF	43.42	48.56	32.32
VW	TOTAL H2O COLLECTED	ML	97.9	94.9	113.0
VWV	VOL H2O VAPOR-STD COND	SCF	4.61	4.47	5.32
PMOS	PERCENT MOISTURE BY VOLUME		9.6	8.4	14.1
MD	MOLE FRACTION DRY GAS		0.904	0.916	0.859
PCO2	PERCENT CO2 BY VOLUME, DRY		11.5	11.8	10.4
PO2	PERCENT O2 BY VOLUME, DRY		7.8	7.8	8.7
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		80.8	80.5	80.9
MWD	MOLECULAR WT-DRY STK GAS		30.15	30.19	30.01
MW	MOLECULAR WT-STK GAS		28.99	29.16	28.32
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.557	0.571	0.854
TS	AVG STACK TEMPERATURE	DEG.F	309.7	294.5	175.8
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.69	28.70	29.58
VS	AVG STACK GAS VELOCITY	FPM	2310	2340	3210
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	321000	335000	641000
QA	ACTUAL STACK FLOWRATE	ACFM	539000	545000	909000
PERI	PERCENT ISOKINETIC		102.3	106.8	105.5
MT	PARTICULATE WT-TOTAL	MG	1194.70	1515.90	96.70
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.425	0.482	0.046
C	PART. LOAD-TTL, STD	LB/DSCF	.00006070	.00006880	.00000659
CAT	PART. LOAD-TTL, STK	GR/ACF	0.252	0.296	0.033
CAW	PART. EMIS-TOTAL	LB/HR	1167.1276	1384.0506	254.0000
E	EMISSION RATE	LB/MMBTU	0.9470	1.0700	0.1110
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.66E-04	3.80E-04	2.56E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.18E-04	2.34E-04	1.80E-05
SCHR	SO2 EMISSION RATE	LB/HR	7051.010	7645.321	983.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.720	5.930	0.430
SCPPM	SO2, CONC., STD	PPM	2210	2290	154
SCPPMA	SO2, CONC., STK	PPM	2000	2100	133

OUTPUT DATE - 08/27/80
TIME - 14:32:46

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS	ESPA04	ESPB04	STK004
	TITLE OF RUN				
	DATE OF RUN		07/20/80	07/20/80	07/20/80
	TIME OF RUN		1230-1454	1230-1448	1235-1424
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0245	1.0355	1.0155
PB	BAROMETRIC PRES, LAB	IN.HG	29.85	29.85	29.85
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.80	29.80	29.63
PM	AVG ORF PRES DROP	IN.H2O	0.621	0.634	0.478
VM	VOL DRY GAS-METER COND	DCF	44.76	45.23	34.49
TM	AVG GAS METER TEMP	DEG.F	118.1	121.7	118.6
VMSTD	VOL DRY GAS-STD COND	DSCF	41.77	42.40	31.69
VW	TOTAL H2O COLLECTED	ML	96.0	68.5	115.7
VWV	VOL H2O VAPOR-STD COND	SCF	4.52	3.22	5.45
PMOS	PERCENT MOISTURE BY VOLUME		9.8	7.1	14.7
MD	MOLE FRACTION DRY GAS		0.902	0.929	0.853
PCO2	PERCENT CO2 BY VOLUME, DRY		12.8	12.1	10.8
PO2	PERCENT O2 BY VOLUME, DRY		8.5	9.0	8.2
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		78.8	78.9	81.0
MWD	MOLECULAR WT-DRY STK GAS		30.38	30.30	30.06
MW	MOLECULAR WT-STK GAS		29.17	29.43	28.29
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.541	0.504	0.855
TS	AVG STACK TEMPERATURE	DEG.F	301.2	280.4	179.2
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.75	28.75	29.57
VS	AVG STACK GAS VELOCITY	FPM	2220	2030	3220
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	311000	302000	636000
QA	ACTUAL STACK FLOWRATE	ACFM	518000	474000	913000
PERI	PERCENT ISOKINETIC		101.3	103.5	104.2
MT	PARTICULATE WT-TOTAL	MG	1080.90	859.40	91.20
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.399	0.327	0.044
C	PART. LOAD-TTL, STD	LB/DSCF	.00005700	.00004680	.00000634
CAT	PART. LOAD-TTL, STK	GR/ACF	0.240	0.208	0.031
CAW	PART. EMIS-TOTAL	LB/HR	1066.0365	847.0000	242.0000
E	EMISSION RATE	LB/MMBTU	0.9440	0.8060	0.1030
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.96E-04	4.10E-04	2.63E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.38E-04	2.61E-04	1.83E-05
SCHR	SO2 EMISSION RATE	LB/HR	7394.102	7437.808	1003.707
SCMM	SO2 EMISSION RATE	LB/MMBTU	6.550	7.080	0.425
SCPPM	SO2, CONC., STD	PPM	2390	2480	159
SCPPMA	SO2, CONC., STK	PPM	2150	2300	135

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OUTPUT DATE - 08/27/80
TIME - 14:33:13

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA05	ESPB05	STK005
	DATE OF RUN		07/21/80	07/21/80	07/21/80
	TIME OF RUN		1030-1243	1030-1233	1030-1219
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0250	1.0350	1.0155
PB	BAROMETRIC PRES, LAB	IN.HG	29.88	29.88	29.88
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.83	29.83	29.66
PM	AVG ORF PRES DROP	IN.H20	0.648	0.613	0.462
VM	VOL DRY GAS-METER COND	DCF	45.96	44.19	33.86
TM	AVG GAS METER TEMP	DEG.F	110.9	116.5	115.2
VMSTD	VOL DRY GAS-STD COND	DSCF	43.50	41.82	31.32
VW	TOTAL H20 COLLECTED	ML	102.5	97.3	115.1
VWV	VOL H20 VAPOR-STD COND	SCF	4.82	4.58	5.42
PMOS	PERCENT MOISTURE BY VOLUME		10.0	9.9	14.7
MD	MOLE FRACTION DRY GAS		0.900	0.901	0.853
PCO2	PERCENT CO2 BY VOLUME, DRY		12.8	13.0	10.7
PO2	PERCENT O2 BY VOLUME, DRY		7.8	8.3	8.1
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		79.5	78.7	81.2
MWD	MOLECULAR WT-DRY STK GAS		30.35	30.41	30.03
MW	MOLECULAR WT-STK GAS		29.12	29.19	28.26
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H20	0.563	0.500	0.855
TS	AVG STACK TEMPERATURE	DEG.F	308.9	305.3	174.3
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.73	28.73	29.60
VS	AVG STACK GAS VELOCITY	FPM	2320	2060	3210
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	322000	287000	638000
QA	ACTUAL STACK FLOWRATE	ACFM	542000	480000	909000
PERI	PERCENT ISOKINETIC		102.1	107.5	102.6
MT	PARTICULATE WT-TOTAL	MG	1397.10	1068.60	90.30
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.496	0.394	0.044
C	PART. LOAD-TTL, STD	LB/DSCF	.00007080	.00005630	.00000635
CAT	PART. LOAD-TTL, STK	GR/ACF	0.294	0.235	0.031
CAW	PART. EMIS-TOTAL	LB/HR	1367.5283	969.0000	243.0000
E	EMISSION RATE	LB/MMBTU	1.1100	0.9200	0.1020
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.87E-04	3.99E-04	2.27E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.30E-04	2.38E-04	1.59E-05
SCHR	SO2 EMISSION RATE	LB/HR	7474.553	6860.159	870.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	6.040	6.510	0.364
SCPPM	SO2, CONC., STD	PPM	2330	2410	137
SCPPMA	SO2, CONC., STK	PPM	2100	2170	117

OUTPUT DATE - 08/27/80
TIME - 14:33:45

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA06	ESPB06	STK006
	DATE OF RUN		07/22/80	07/22/80	07/22/80
	TIME OF RUN		1104-1338	1109-1416	1035-1216
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0245	1.0350	1.0145
PB	BAROMETRIC PRES, LAB	IN.HG	29.79	29.79	29.79
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.74	29.74	29.57
PM	AVG ORF PRES DROP	IN.H2O	0.607	0.613	0.440
VM	VOL DRY GAS-METER COND	DCF	44.24	43.12	32.82
TM	AVG GAS METER TEMP	DEG.F	89.4	98.7	103.2
VMSTD	VOL DRY GAS-STD COND	DSCF	43.36	41.98	30.88
VW	TOTAL H2O COLLECTED	ML	108.6	101.8	111.7
VWV	VOL H2O VAPOR-STD COND	SCF	5.11	4.79	5.26
PMOS	PERCENT MOISTURE BY VOLUME		10.5	10.2	14.5
MD	MOLE FRACTION DRY GAS		0.895	0.898	0.855
PCO2	PERCENT CO2 BY VOLUME, DRY		11.1	11.5	11.1
PO2	PERCENT O2 BY VOLUME, DRY		8.1	8.0	7.7
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		80.8	80.5	81.2
MWD	MOLECULAR WT-DRY STK GAS		30.10	30.16	30.08
MW	MOLECULAR WT-STK GAS		28.83	28.92	28.32
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.546	0.523	0.851
TS	AVG STACK TEMPERATURE	DEG.F	298.0	300.8	175.9
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.67	28.67	29.51
VS	AVG STACK GAS VELOCITY	FPM	2250	2160	3200
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	314000	301000	635000
QA	ACTUAL STACK FLOWRATE	ACFM	526000	504000	907000
PERI	PERCENT ISOKINETIC		104.3	102.8	101.8
MT	PARTICULATE WT-TOTAL	MG	1212.20	1788.20	91.30
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.431	0.657	0.046
C	PART. LOAD-TTL, STD	LB/DSCF	.00006160	.00009390	.00000651
CAT	PART. LOAD-TTL, STK	GR/ACF	0.258	0.392	0.032
CAW	PART. EMIS-TOTAL	LB/HR	1161.2044	1695.5126	248.0000
E	EMISSION RATE	LB/MMBTU	0.9900	1.4900	0.1020
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.90E-04	4.11E-04	4.11E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.33E-04	2.45E-04	2.88E-05
SCHR	SO2 EMISSION RATE	LB/HR	7343.135	7420.226	1564.931
SCMM	SO2 EMISSION RATE	LB/MMBTU	6.260	6.540	0.641
SCPPM	SO2, CONC., STD	PPM	2350	2480	248
SCPPMA	SO2, CONC., STK	PPM	2100	2230	212

OUTPUT DATE - 08/27/80
TIME - 14:34:15

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS	ESPA07	ESPB07	STK007
	TITLE OF RUN				
	DATE OF RUN		07/23/80	07/23/80	07/23/80
	TIME OF RUN		1620-1832	1620-1823	1620-1814
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0230	1.0320	1.0115
PB	BAROMETRIC PRES, LAB	IN.HG	29.80	29.80	29.80
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.75	29.75	29.58
PM	AVG ORF PRES DROP	IN.H2O	0.535	0.450	0.364
VM	VOL DRY GAS-METER COND	DCF	41.87	39.70	29.43
TM	AVG GAS METER TEMP	DEG.F	94.7	111.8	102.2
VMSTD	VOL DRY GAS-STD COND	DSCF	40.58	37.65	27.66
VW	TOTAL H2O COLLECTED	ML	90.0	80.8	93.2
VWV	VOL H2O VAPOR-STD COND	SCF	4.24	3.80	4.39
PMOS	PERCENT MOISTURE BY VOLUME		9.5	9.2	13.7
MD	MOLE FRACTION DRY GAS		0.905	0.908	0.863
PCO2	PERCENT CO2 BY VOLUME, DRY		10.7	11.5	10.5
PO2	PERCENT O2 BY VOLUME, DRY		7.3	7.8	7.8
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.1
PN2	PERCENT N2 BY VOLUME, DRY		82.0	80.7	81.5
MWD	MOLECULAR WT-DRY STK GAS		30.00	30.15	30.00
MW	MOLECULAR WT-STK GAS		28.87	29.04	28.36
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.516	0.449	0.785
TS	AVG STACK TEMPERATURE	DEG.F	304.5	284.6	170.7
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.80	28.82	29.52
VS	AVG STACK GAS VELOCITY	FPM	2130	1820	2930
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	300000	264000	593000
QA	ACTUAL STACK FLOWRATE	ACFM	498000	426000	832000
PERI	PERCENT ISOKINETIC		102.4	105.0	97.6
MT	PARTICULATE WT-TOTAL	MG	718.50	921.70	81.10
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.273	0.378	0.045
C	PART. LOAD-TTL, STD	LB/DSCF	.00003900	.00005400	.00000646
CAT	PART. LOAD-TTL, STK	GR/ACF	0.164	0.234	0.032
CAW	PART. EMIS-TOTAL	LB/HR	701.0000	856.0000	230.0000
E	EMISSION RATE	LB/MMBTU	0.5890	0.8470	0.1020
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.38E-04	3.37E-04	2.27E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.03E-04	2.09E-04	1.62E-05
SCHR	SO2 EMISSION RATE	LB/HR	6073.866	5338.111	807.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.100	5.290	0.357
SCPPM	SO2, CONC., STD	PPM	2040	2030	137
SCPPMA	SO2, CONC., STK	PPM	1850	1840	118

OUTPUT DATE - 08/27/80
TIME - 14:34:47

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA08	ESPB08	STK008
	DATE OF RUN		07/24/80	07/24/80	07/24/80
	TIME OF RUN		1100-1310	1100-1302	1100-1244
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0245	1.0340	1.0130
PB	BAROMETRIC PRES, LAB	IN.HG	29.82	29.82	29.82
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.77	29.77	29.69
PM	AVG ORF PRES DROP	IN.H2O	0.607	0.563	0.397
VM	VOL DRY GAS-METER COND	DCF	44.31	42.69	30.98
TM	AVG GAS METER TEMP	DEG.F	93.5	107.9	100.3
VMSTD	VOL DRY GAS-STD COND	DSCF	43.14	40.89	29.28
VW	TOTAL H2O COLLECTED	ML	107.5	99.3	109.1
VWV	VOL H2O VAPOR-STD COND	SCF	5.06	4.67	5.14
PMOS	PERCENT MOISTURE BY VOLUME		10.5	10.3	14.9
MD	MOLE FRACTION DRY GAS		0.895	0.897	0.851
PCO2	PERCENT CO2 BY VOLUME, DRY		11.2	11.5	10.8
PO2	PERCENT O2 BY VOLUME, DRY		7.3	7.2	7.7
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		81.5	81.3	81.4
MWD	MOLECULAR WT-DRY STK GAS		30.08	30.13	30.04
MW	MOLECULAR WT-STK GAS		28.81	28.88	28.25
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.549	0.512	0.821
TS	AVG STACK TEMPERATURE	DEG.F	309.3	294.9	167.8
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.72	28.65	29.55
VS	AVG STACK GAS VELOCITY	FPM	2280	2110	3070
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	314000	296000	614000
QA	ACTUAL STACK FLOWRATE	ACFM	532000	492000	869000
PERI	PERCENT ISOKINETIC		103.9	102.0	99.7
MT	PARTICULATE WT-TOTAL	MG	808.50	805.60	76.30
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.289	0.304	0.040
C	PART. LOAD-TTL, STD	LB/DSCF	.00004130	.00004340	.00000574
CAT	PART. LOAD-TTL, STK	GR/ACF	0.170	0.183	0.028
CAW	PART. EMIS-TOTAL	LB/HR	778.0000	770.0000	212.0000
E	EMISSION RATE	LB/MMBTU	0.6250	0.6490	0.0893
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	3.52E-04	3.47E-04	2.15E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.08E-04	2.09E-04	1.52E-05
SCHR	SO2 EMISSION RATE	LB/HR	6629.854	6158.371	794.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.330	5.190	0.335
SCPPM	SO2, CONC., STD	PPM	2120	2100	130
SCPPMA	SO2, CONC., STK	PPM	1900	1880	111

OUTPUT DATE - 08/27/80
TIME - 14:35:21

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS	ESPA09	ESPB09	STK009
	TITLE OF RUN				
	DATE OF RUN		07/25/80	07/25/80	07/25/80
	TIME OF RUN		1145-1344	1145-1350	1145-1321
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0240	1.0338	1.0135
PB	BAROMETRIC PRES, LAB	IN.HG	29.84	29.84	29.84
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.79	29.79	29.62
PM	AVG ORF PRES DROP	IN.H2O	0.590	0.542	0.437
VM	VOL DRY GAS-METER COND	DCF	43.53	42.35	33.01
TM	AVG GAS METER TEMP	DEG.F	98.9	113.6	104.7
VMSTD	VOL DRY GAS-STD COND	DSCF	41.99	40.17	31.00
VW	TOTAL H2O COLLECTED	ML	95.9	87.2	110.0
VWV	VOL H2O VAPOR-STD COND	SCF	4.51	4.10	5.18
PMOS	PERCENT MOISTURE BY VOLUME		9.7	9.3	14.3
MD	MOLE FRACTION DRY GAS		0.903	0.907	0.857
PCO2	PERCENT CO2 BY VOLUME, DRY		11.1	9.3	10.7
PO2	PERCENT O2 BY VOLUME, DRY		8.5	9.3	7.9
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.0
PN2	PERCENT N2 BY VOLUME, DRY		80.4	81.5	81.4
MWD	MOLECULAR WT-DRY STK GAS		30.12	29.85	30.03
MW	MOLECULAR WT-STK GAS		28.95	28.75	28.31
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.541	0.492	0.860
TS	AVG STACK TEMPERATURE	DEG.F	300.4	294.8	174.7
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.74	28.74	29.57
VS	AVG STACK GAS VELOCITY	FPM	2230	2030	3220
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	313000	288000	644000
QA	ACTUAL STACK FLOWRATE	ACFM	521000	473000	914000
PERI	PERCENT ISOKINETIC		101.2	102.7	100.7
MT	PARTICULATE WT-TOTAL	MG	907.40	1242.00	98.40
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.334	0.477	0.049
C	PART. LOAD-TTL, STD	LB/DSCF	.00004760	.00006820	.00000699
CAT	PART. LOAD-TTL, STK	GR/ACF	0.201	0.291	0.035
CAW	PART. EMIS-TOTAL	LB/HR	896.0000	1178.8716	270.0000
E	EMISSION RATE	LB/MMBTU	0.7890	1.2000	0.1100
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	4.19E-04	4.06E-04	2.21E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.52E-04	2.47E-04	1.56E-05
SCHR	SO2 EMISSION RATE	LB/HR	7876.590	7019.656	854.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	6.930	7.150	0.349
SCPPM	SO2, CONC., STD	PPM	2530	2450	133
SCPPMA	SO2, CONC., STK	PPM	2280	2220	114

OUTPUT DATE - 08/27/80
TIME - 15:23:52

KENVIRONS, INC.
PROJECT NO. 80019

BECHTEL 1980 DEMO-ACCEPTANCE TEST - LG&E CANE RUN NO.6

PARTICULATE AND SO2 EMISSION DATA

NAME	DESCRIPTION	UNITS			
	TITLE OF RUN		ESPA10	ESPB10	STK010
	DATE OF RUN		07/26/80	07/26/80	07/26/80
	TIME OF RUN		0930-1138	0930-1129	0935-1116
DN	PROBE TIP DIAMETER	IN	0.238	0.241	0.172
TT	NET TIME OF RUN	MIN	100.0	100.0	84.0
NP	NUMBER OF SAMPLING POINTS		50	50	28
Y	DRY GAS METER CALIB. FACTOR		1.0235	1.0340	1.0130
PB	BAROMETRIC PRES, LAB	IN.HG	29.78	29.78	29.78
PBC	BAROMETRIC PRES, CONSOLE	IN.HG	29.73	29.73	29.56
PM	AVG ORF PRES DROP	IN.H2O	0.561	0.550	0.398
VM	VOL DRY GAS-METER COND	DCF	42.84	41.53	31.38
TM	AVG GAS METER TEMP	DEG.F	99.4	108.3	105.1
VMSTD	VOL DRY GAS-STD COND	DSCF	41.18	39.70	29.37
VW	TOTAL H2O COLLECTED	ML	92.9	86.5	100.6
VWV	VOL H2O VAPOR-STD COND	SCF	4.37	4.07	4.74
PMOS	PERCENT MOISTURE BY VOLUME		9.6	9.3	13.9
MD	MOLE FRACTION DRY GAS		0.904	0.907	0.861
PCO2	PERCENT CO2 BY VOLUME, DRY		11.5	12.8	10.7
PO2	PERCENT O2 BY VOLUME, DRY		7.0	7.3	7.9
PCO	PERCENT CO BY VOLUME, DRY		0.0	0.0	0.1
PN2	PERCENT N2 BY VOLUME, DRY		81.5	80.0	81.3
MWD	MOLECULAR WT-DRY STK GAS		30.12	30.33	30.03
MW	MOLECULAR WT-STK GAS		28.96	29.18	28.36
CP	PITOT TUBE COEFFICIENT		0.840	0.840	0.840
DPS	AVG SQRT STK VEL HEAD	IN.H2O	0.523	0.499	0.823
TS	AVG STACK TEMPERATURE	DEG.F	300.4	291.9	172.5
PS	STACK PRESSURE, ABSOLUTE	IN.HG	28.57	28.64	29.50
VS	AVG STACK GAS VELOCITY	FPM	2160	2040	3080
AS	STACK AREA	IN2	33618	33618	40828
QS	STK FLOWRATE, DRY, STD	DSCFM	302000	290000	620000
QA	ACTUAL STACK FLOWRATE	ACFM	505000	476000	874000
PERI	PERCENT ISOKINETIC		102.9	100.9	99.2
MT	PARTICULATE WT-TOTAL	MG	1216.70	710.20	86.00
CAN	PART. LOAD-TTL, STD	GR/DSCF	0.456	0.276	0.045
C	PART. LOAD-TTL, STD	LB/DSCF	.00006510	.00003940	.00000645
CAT	PART. LOAD-TTL, STK	GR/ACF	0.273	0.168	0.032
CAW	PART. EMIS-TOTAL	LB/HR	1181.9692	686.0000	240.0000
E	EMISSION RATE	LB/MMBTU	0.9620	0.5930	0.1020
F	F FACTOR	DSCF/MMBTU	9820	9820	9820
SCONC	SO2, CONC., STD	LB/DSCF	4.00E-04	3.92E-04	2.27E-05
SCONCA	SO2, CONC., STK	LB/ACF	2.40E-04	2.39E-04	1.61E-05
SCHR	SO2 EMISSION RATE	LB/HR	7256.073	6816.098	845.000
SCMM	SO2 EMISSION RATE	LB/MMBTU	5.900	5.890	0.359
SCPPM	SO2, CONC., STD	PPM	2410	2360	137
SCPPMA	SO2, CONC., STK	PPM	2180	2140	118

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